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DESIGN OF FILTERS FOR

ADVANCED WASTEWATER TREATMENT

by

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DESIGN OF FILTERS FOR ADVANCED WASTEWATER TREATMENT

INTRODUCTION

If water containing suspended solids is passed through a layer of porous media, some of the suspended and colloidal material are partially removed. This process is called FILTRATION and its efficiency and cost is a function of:

- the concentration and characteristics of the solids in suspension (particle size distribution, surface characteristics, organic vs. inorganic, etc.),
- the characteristics of the filter media and other filtering aids used (particle size distribution, surface characteristics, etc.),
- the characteristics of the solids in solution in the water filtered, and
- the characteristics of the filter and its method of operation.

The criteria of importance which must be considered in design involve finding:

the OPERATIONAL OPTIMUM filter design characteristics, and

the ECONOMIC OPTIMUM filter design.

During a filter run, the head loss across the filter media will increase due to the accumulation of solids within the filter media. When this head reaches the limit set by the hyraulic conditions of the design, the filter run must be stopped. Figure 1 indicates that as the head loss increases during a filter run, the filtrate quality also changes, tending to rise in value as time proceeds. Although the filter could be designed to produce a satisfactory filtrate quality during the early stages of the run, there will come a time when the filtrate quality will become unsatisfactory and the filter run will have to be terminated due to solids breakthrough above the maximum permissible critical concentration of suspended

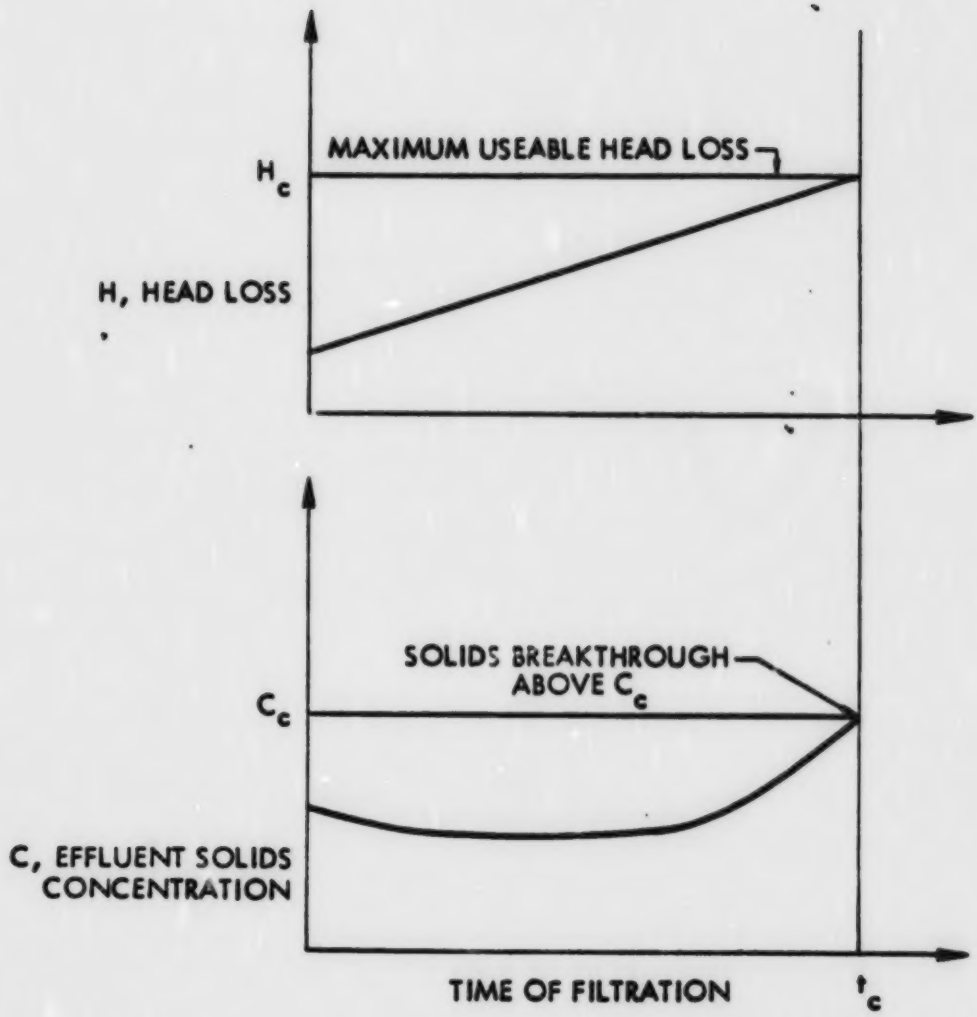


Fig. 1. Operational Optimum Filter Design.

solids, (Cc). A filter OPERATIONAL OPTIMUM condition occurs when both the head loss and effluent quality reach their respective critical values (Hc and Cc) at the same time.

To achieve an OPERATIONAL OPTIMUM, many alternative designs are possible which can produce "equivalent performance." Two or more filters may be said to provide "equivalent performance" when they produce the same quality and quantity of filtered water from the same water source during the same time period. Of the equivalent performance filters, however, only one will produce water at the least total cost per thousand gallons. Current trends indicate that we are approaching the time when we will be able to design filters to provide both OPERATIONAL and ECONOMIC (Least Cost) OPTIMUMS.

Operational optimum design requires consideration only of the:

- Media sizes,
- Media depths,
- Filtration rate, and
- Hydraulic conditions affecting the head loss.

In a least cost optimum design, the cost effects of the filter structure, hydraulic appurtenances, energy, labor and maintenance requirements, and the size of the total filter plant also have to be considered. Ives (1) and Huang and Baumann (2) have both recently considered in detail the data needed for design of OPERATIONAL and ECONOMIC or LEAST COST OPTIMUM designs for granular media filters suitable for water and wastewater treatment applications. To accomplish this, three elements are necessary:

- A method (mathematical model) to predict the performance of the filters to determine combinations of the filter design variables which will provide the same quality of effluent under operational optimum conditions.
- A filtration system whose first cost, operating cost, and maintenance cost can be predicted with reasonable accuracy.

- A computer program which can provide a list of operational optimum filter designs and predict the one design which will provide the treated water at least cost.

To date, no such mathematical model -- or even filter operating experience with wastewater -- is available for consideration in tertiary filtration applications. Accordingly, it would appear that prior to detail design of filters for wastewater treatment the required operational optimum least cost mathematical model should be developed by:

- Running pilot plant studies using a water pretreated as proposed in the tertiary filtration application.
- Building the complete treatment plant based on pilot plant design with enough filter flexibility for perfecting the model using full-scale plant filters and the actual waste to be treated.

The analysis of the data collected to produce a satisfactory model could be accomplished using the techniques of Hsuing-Cleasby (3). Until such data are available, the design requirements of tertiary filtration plants must be based on the wastewater filtration experience of others.

APPLICATIONS FOR WASTEWATER FILTRATION

Most of the published information useful in the design of wastewater filters has been drawn from experiences in water filtration. Filters have been used in potable water production for the removal of solids present in surface waters either with or without pretreatment by coagulation and sedimentation, for the removal of precipitates resulting from lime or lime-soda ash softening, and for the removal of iron and/or manganese found in many underground water supplies. The design and operating experiences in these applications has been carried over into wastewater applications, sometimes disastrously.

In wastewater treatment, filtration is used for:

- Removal of residual biological floc in settled effluents from secondary treatment by trickling filters or activated sludge processes.
- Removal of precipitates resulting from alum, iron, or lime precipitation of phosphates in secondary treatments from trickling filters or activated sludge processes.
- Removal of solids remaining after the chemical coagulation of wastewaters in physicochemical advanced waste treatment processes, i.e. lime treatment of raw wastewater prior to carbon adsorption for soluble carbon removal.

Filters may be used as the final process of advanced waste treatment (polishing secondary effluents) or as an intermediate process to prepare wastewater for further treatment (prior to downflow carbon adsorption columns, or clinoptilolite ammonia exchange columns, or reverse osmosis treatment, for example). In either case, the design of the filters required should be based on the desired quality of the filtrate and achieving this quality will require evaluation of the flow characteristics and solids characteristics (concentration, particle size distribution, and surface properties) of the water to be filtered. Unfortunately, too many wastewater filters are provided

without evaluation of what they can and should do, but merely because they are expected to improve effluent quality. As a result, it is still nearly impossible to design to achieve a given effluent quality and reliance is placed on designing for a given filter operating experience (a given length of run at the design filtration rate) with the acceptance of the effluent quality that results. Appendix A represents a review of published papers describing use of granular filters for wastewater treatment.

FILTER DESIGN CONSIDERATIONS

General

The design of a filter for a given application requires consideration of the following:

- Selection of the filter configuration,
- Selection of the method of filter operation,
- Selection of the filter media, sizes and depths,
- Selection of the backwashing requirements,
- Selection of the terminal head loss, ft. of water, and
- Selection of the filtration rate, gpm/sq.ft.

Since the capital cost of a filter is chiefly a function of the area of filter provided, the use of a high filtration rate is usually preferred. In general, the filter design should seek to maximize the net water production per square foot of filter consistent with filter operating feasibility. A most useful relationship between net water production and run lengths obtained at different filtration rates is shown in Figure 2. This figure was constructed by calculating the net useful water production at various filtration rates when run lengths (hours) of 1, 2, 3, 5, 10, 20, 30, 50, and infinite duration are obtained. Backwash is assumed to require 30 minutes, including a water wash of 5 minutes at 20 gpm/sq.ft., preceded by 3 minutes of air wash at a rate of 3 CFM/sq.ft. Figure 2 shows that there exists an upper limit of net water production at each flow rate. The maximum net water production which can be obtained in a day is 2,880, 5,760, and 8,640 gallons per day per sq. ft. at filtration rates of 2, 4, and 6 gpm/sq.ft., respectively. Assume that a net water production of 3,500 gpd/sq.ft. is desirable. This can be obtained at a filtration rate of 2.45 gpm/sq.ft.

if an infinite run length can be obtained. At 2.5 gpm/sq.ft., a 50 hour run is needed. At 3 gpm/sq.ft., a run length of 5 hours is needed. With a one hour run length, a 5.3 gpm/sq.ft. filtration rate could be used. In summary, Figure 2 indicates that any run length greater than one hour at 5.3 gpm/sq.ft. would produce more net water production than an infinite run length at 2.45 gpm/sq.ft. This means that the design combinations used can be judged, in part, on the basis of economic filter design.

Filter Configurations

Figure 3 shows several filter configurations which have been used extensively in the past few years in water filtration. The granular filter originally was used for potable water filtration with a single media (effective size = 0.5 mm, 1.5-1.8 uniformity coefficient) and downward filtration (first through the finest media which collected at the top of the filter) at low filtration rates (2 gpm/sq.ft) (Figure 3a). In recent years, design trends are to use of higher filtration rates (3 to 6 gpm/sq.ft.), deeper media, and coarser media. To better distribute solids within the filter media, several filter configurations are in use:

- (Fig. 3b) Using upflow filtration through a relatively deep, coarse filter media.
- (Fig. 3c) Using a filtered water collection device within the filter media and bringing water in from both the bottom and top of the filter media.
- (Fig. 3d, e) Using dual or multi-media with downward flow as in conventional filters.

British practice in tertiary treatment tends to use the Fig. 3b configuration (4). This configuration has several advantages:

- Filtration can take place from coarse to fine media using a single media for better solids distribution. Therefore, a coarser effective size with a larger uniformity coefficient media can be used.

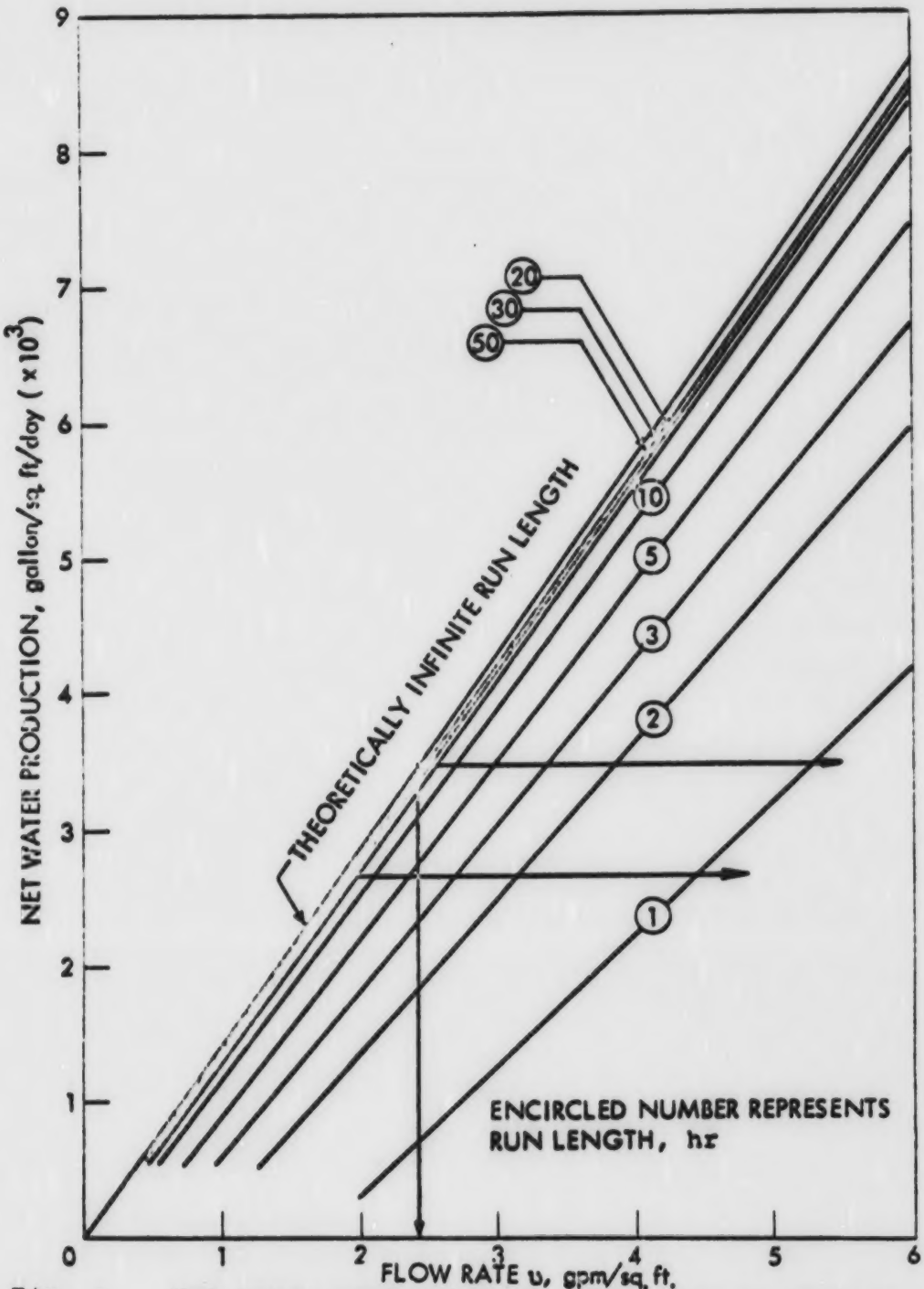
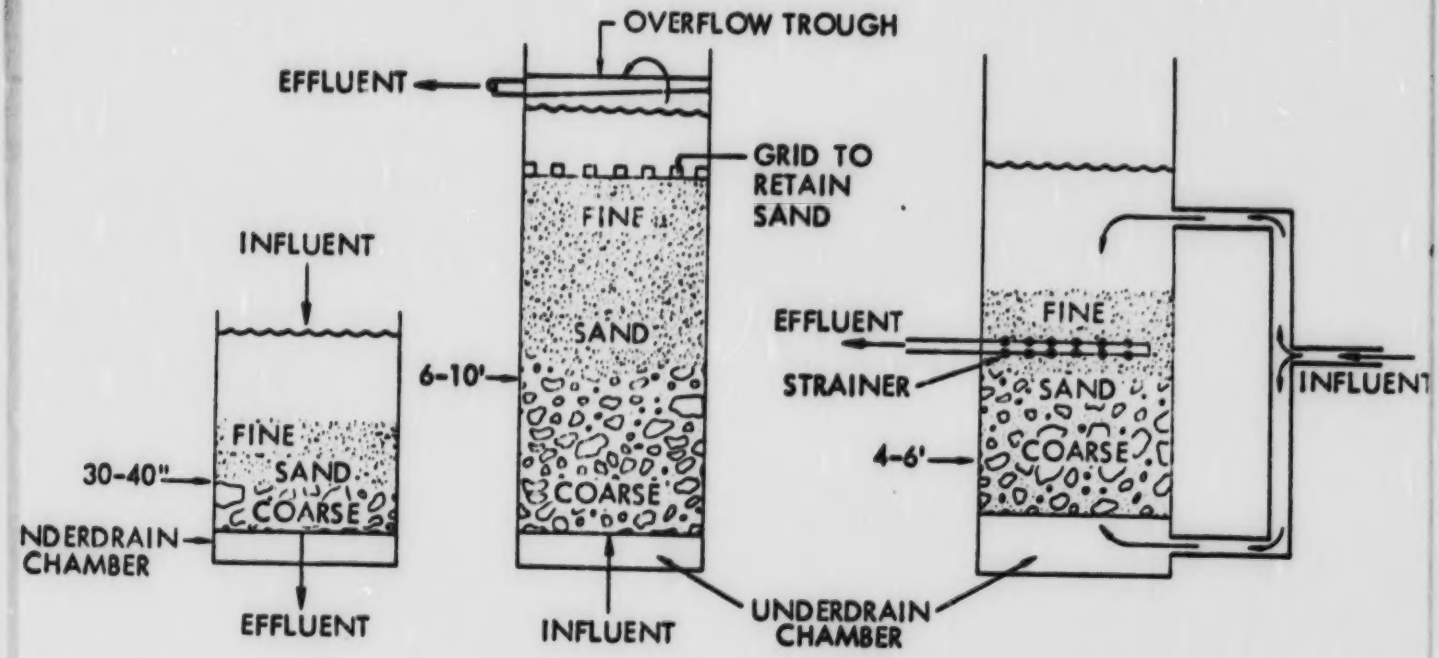
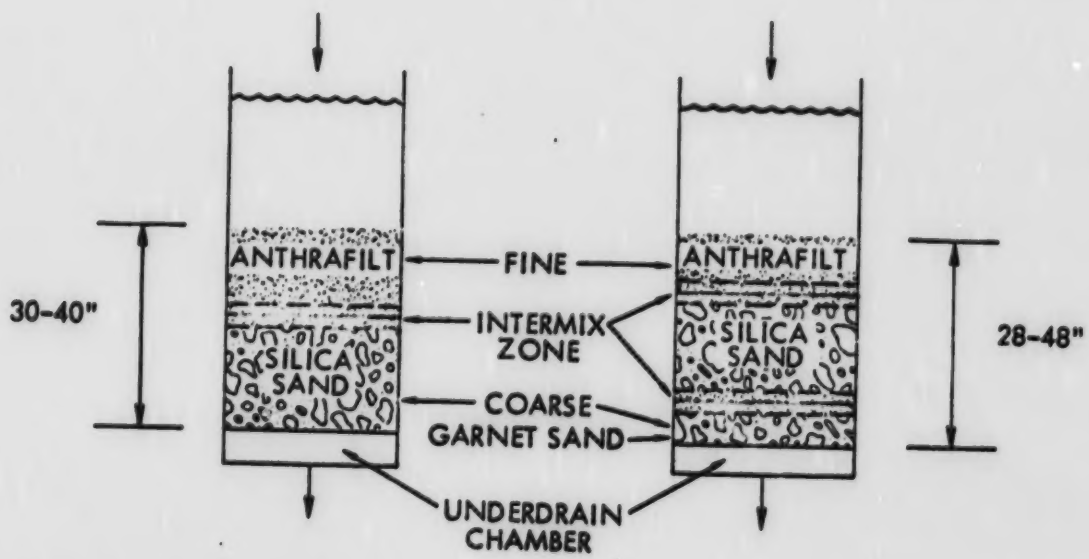


Fig. 2. Net water production vs. flow rate at various run lengths (30 min. backwashing period assumed)



), CONVENTIONAL FILTER (b), UPFLOW FILTER (c), BI-FLOW FILTER

SINGLE MEDIA FILTERS



(d), DUAL MEDIA FILTERS (e), MULTI-MEDIA FILTERS

Fig. 3. Filter Configurations.

- Effective backwash time is less since drain-down time is significantly less.
- Raw water is used for backwashing reducing somewhat the amount of water which must be filtered twice.
- There is no necessity for having high walls on the filter tank in order to build up the necessary static head required for filtration.

The chief disadvantages of this configuration are:

- The need for a bar screen configuration within the sand near the surface to retain the media in place against the upward force exerted during filtration.
- There is a tendency for the media retaining nozzles used to deliver raw water to the filters to clog when suspended solids in the filter are high.

American practice tends to the use of two or three media as in Figs. 3d and e. Usually, a coarse anthrafilt (Sp. Gr. = 1.4-1.6) is used on top of a finer silica sand (Sp. Gr. = 2.65). Occasionally, a still finer garnet sand (Sp. Gr. = 3.5±) is used at the bottom. Dual or multi-media filters have the following advantages:

- Filter design follows current American practice and permits production of the desired quality of effluent with reasonable run lengths.
- No filter media restraining grids or underdrain clogging problems are encountered.

They also have the following disadvantages:

- Dual or multi-media filters require relatively deep filter cells to provide the required filtering head without creating negative head conditions in the filter.
- Filtering down time in preparation for backwashing is significant and extends the out-of-service time for backwashing to about 34 minutes at a filtration rate of 3 gpm/sq.ft. and 24 minutes at 6 gpm/sq.ft.
- Use of dual of multi-media filters creates a need for more care in backwash design to prevent loss of media or excessive intermixing of the media.

Although both Fig. 3b and Fig. 3d and 3e filter configurations have applications in tertiary wastewater filtration, only the dual-media filter configuration will be considered hereafter in this paper.

Methods of Operation

In any filtration operation, the rate of flow through a filter may be expressed as:

$$\text{rate of flow} = \frac{\text{driving force}}{\text{filter resistance}} . \qquad (1)$$

The rate of flow through a water filter is usually expressed in units of gpm/sq.ft. The driving force refers to the pressure drop across the filter which is available to force the water through the filter. At the start of a filter run, the filter is clean and the driving force need only overcome the resistance of the clean filter medium. As filtration continues, the suspended solids removed by the filter collect on the filter surface or in the filter medium or both and the driving force must overcome the combined resistance of the filter medium and the solids removed by the filter. With sand and multimedia filters, the driving force across the filter is usually limited to 5 to 15 ft. of water with a normal value of 8 to 10 ft.

The filter resistance refers to the resistance of the filter medium and the solids removed by the filter medium to the passage of water. The filter resistance increases during a filter run because of the accumulation of the solids removed by the filter. The filter resistance also increases as the pressure drop across the cake increases because the solids already removed compress and become more resistant to flow. Hence, as the filter resistance increases, the driving force across the filter must increase proportionally to maintain a constant rate of flow.

There are three basic methods of operating filters that differ primarily

in the way that the driving force is applied across the filter. These methods are referred to as constant-pressure filtration, constant-rate filtration, and variable declining-rate filtration.

Constant-Pressure Filtration. In true constant-pressure filtration, the total available driving force is applied across the filter throughout the filter run. At the beginning of the filter run, the filter resistance is low and the rate of filtration is very high. (High driving force $\frac{\Delta}{\rho}$ low filter resistance = high rate of flow). As the filter clogs with solids, filter resistance increases, and since the driving force remains constant, the flow rate decreases. This method of operation is represented by Curve "A" in Fig. 4.

Constant-Rate Filtration. The constant-pressure method of filtration is seldom used in water or wastewater filtration practice because it requires a relatively large volume of water storage on the upstream or downstream side of the filter. Current practice, therefore, has tended to the use of constant-rate or declining-rate filtration. In constant-rate filtration, a constant pressure is supplied to the filter system, but the filter resistance is then held constant across the filter by the action of a manually-operated or automatic flow control valve. This method of operation is represented by Curve "B" in Fig.4. At the beginning of the filter run, the filter is clean and has little resistance. If the full driving force were applied across the filter only, the flow rate would be very high. To maintain a constant flow rate, some of the available driving force is consumed by the flow control valve, Fig. 5. At the start of the filter run, the flow control valve is nearly closed in providing the additional resistance needed to keep the flow rate at the level desired. As filtration

continues, the filter becomes clogged with solids, and the flow control valve is opened slowly. When the flow control valve is fully open, the constant-rate run must be terminated, since any further increase in filter resistance will not be balanced by a corresponding decrease in the resistance of the flow control valve. Thus, the ratio of driving force to filter resistance (Eq. (1)) will decrease and flow rate also decreases. The disadvantages of effluent control constant-rate operation include:

- The initial and operating costs of the fairly complex rate-control system are high, and
- The filtered water quality is not as good using gravity, granular media filters as that obtained using declining-rate filter operation.[5]

Rate control for constant-rate gravity granular filters has been achieved for many years by such effluent control systems. The valves used as rate controllers frequently do not function properly and require an excessive amount of maintenance. A number of other alternative methods of flow control are coming into use that will supplant the effluent flow control valve[5]. For example, some plants have been constructed so as to split the flow nearly equally (<u>influent</u> flow splitting) to all the operating filters, usually by means of an influent weir box on each filter. A schematic diagram of such a filter is shown in Fig. 6. The advantages of this system include:

- Constant-rate filtration is achieved <u>without</u> rate controllers if the total plant flow remains constant,
- When a filter is taken out of service for backwashing or returned to service after backwashing, the water level gradually rises or lowers in the operating filters until sufficient head is achieved to handle the flow. Thus, the rate changes are made slowly and smoothly without the abrupt effects associated with automatic or manual control equipment. This causes the least harmful effect to filtered water quality, [8]

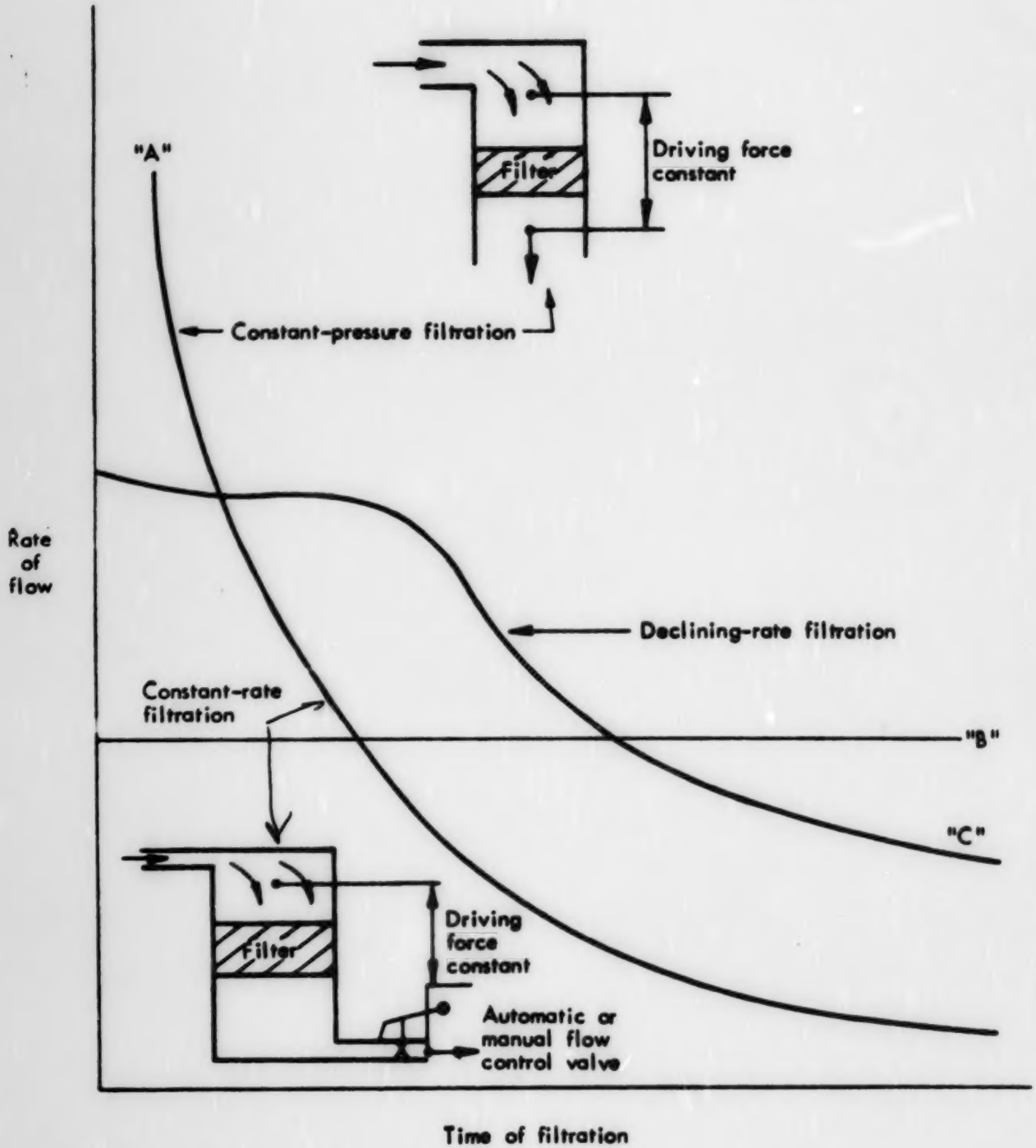


Fig. 4. Methods of Filter Operation.

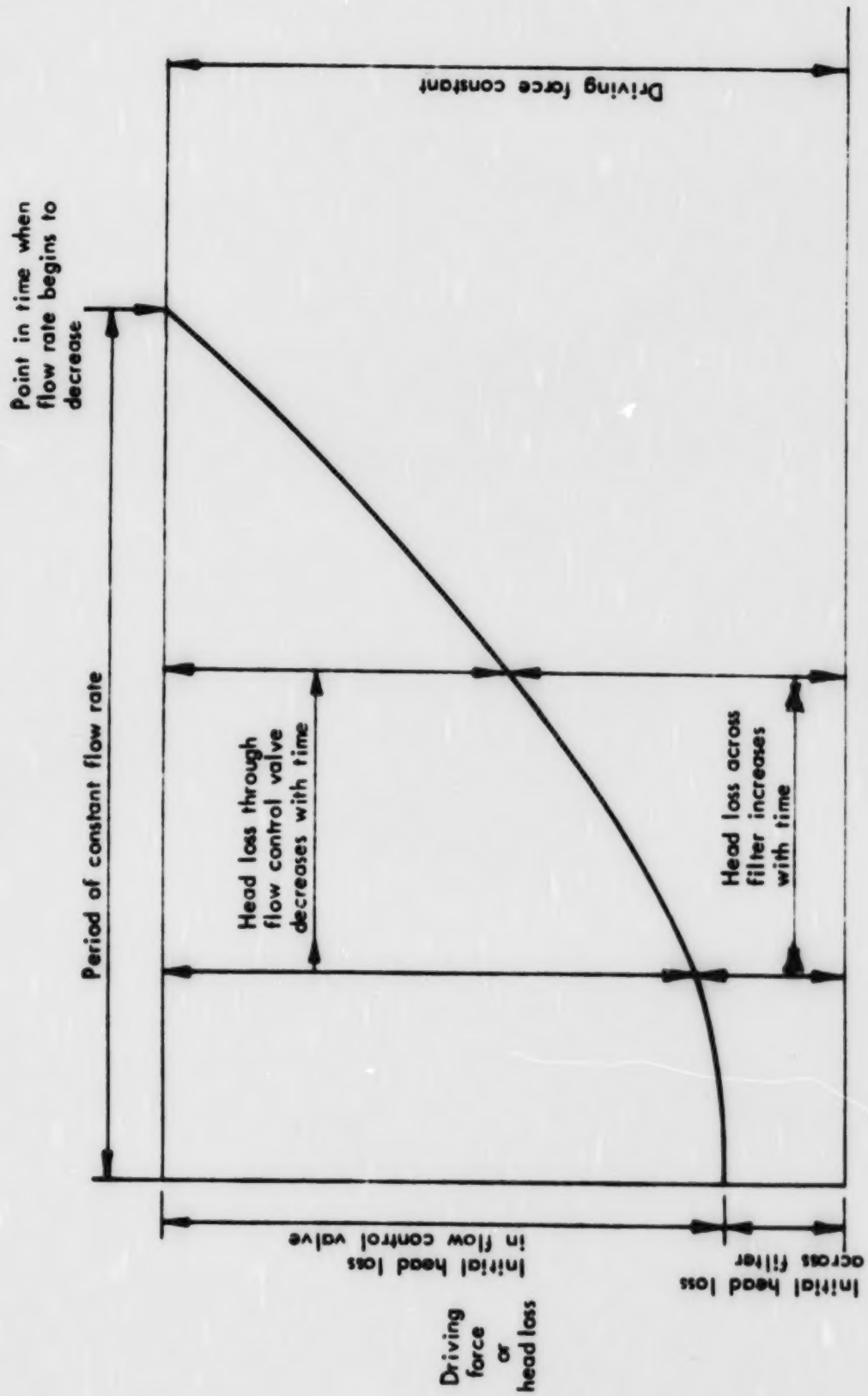

Driving force constant
Point in time when flow rate begins to decrease
Period of constant flow rate
Head loss through flow control valve decreases with time
Head loss across filter increases with time
Initial head loss in flow control valve
Initial head loss across filter
Driving force or head loss
Time of filtration

Fig. 5. Constant-rate Filter Operation

- The head loss for a particular filter is evident to the operator by the water level in the filter box and can be read on a simple staff gage attached to the filter box wall, rather than by elaborate automatic remote reading devices. When the water reaches a desired maximum level (the desired head loss), backwashing of that filter is required. Effluent flow meters for each filter are not essential sin : the flow per filter is the total plant flow divided by the number of filters in service.

- The effluent control weir must be located above the sand to prevent accidental dewatering of the filter bed. This arrangement eliminates completely the possibility of negative head in the filter and the well-known and undesirable problems which sometimes result from it.

Appendix B describes the conditions necessary for the creation of negative head conditions in a filter and the problems created when it exists.

The only disadvantage of the influent-flow splitting system is the additional depth of filter box required. The depth of filter box must be increased 5 to 6 ft. over conventional box depths to provide a reasonable available head loss. Figure 7 shows* an influent flow-splitting filter in both filter and backwashing cycles. In this case, the effluent weir is set above the level of the wash trough so that when one filter is down for backwashing, the flow from the other three filters automatically is used for backwashing. Unfortunately, with the coarse media used in wastewater filtration, the other three filters cannot provide enough water for backwashing one filter and this scheme is not effective.

Declining-Rate Filtration. It is also possible to operate a filter between constant-pressure and constant-rate controls, an operation frequently referred to as declining-rate filtration.

---

* This type of filter is manufactured by the General Filter Co. of Ames, Iowa, under the trade name "CenTROL filter" (central filter control of gravity multicells) and by the Fuller Co., General American Transportation Corp., Tucson, Ariz., under the trade name "Greenleaf filter."

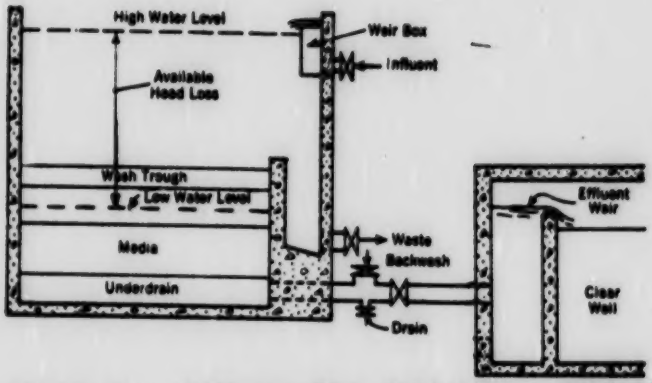


Figure 6a. Influent Flow Splitting - Typical filter and clear well arrangement.

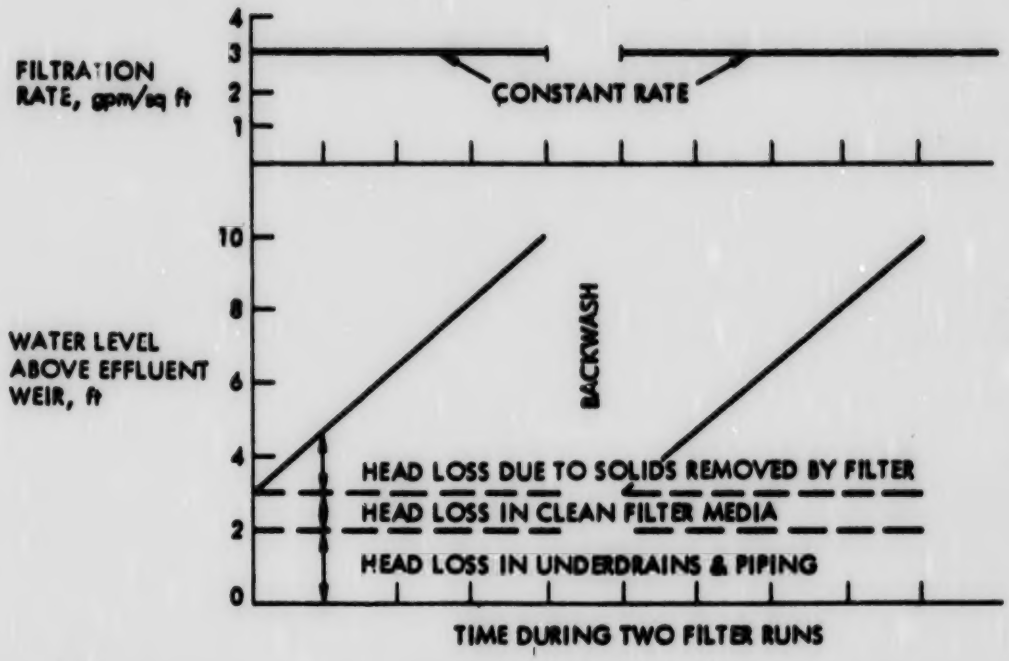
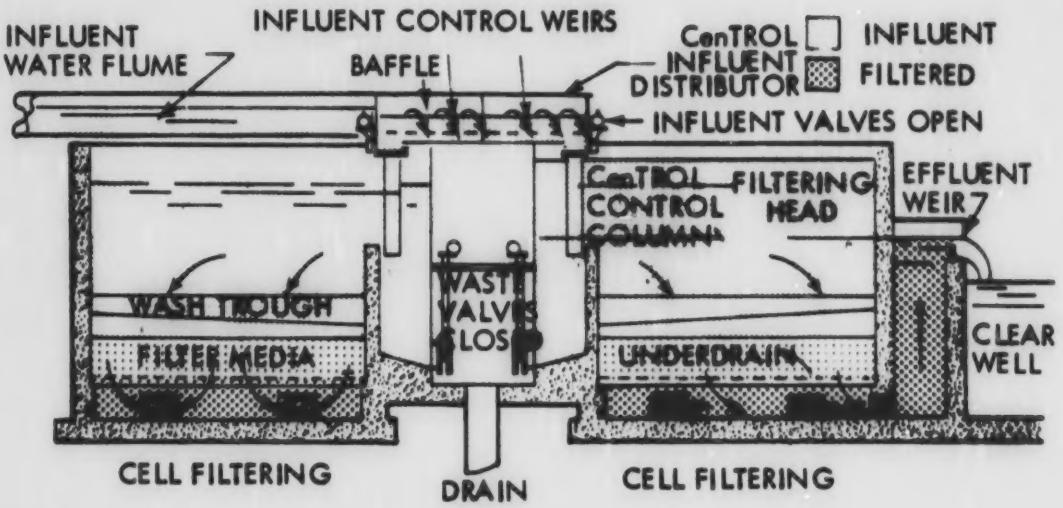
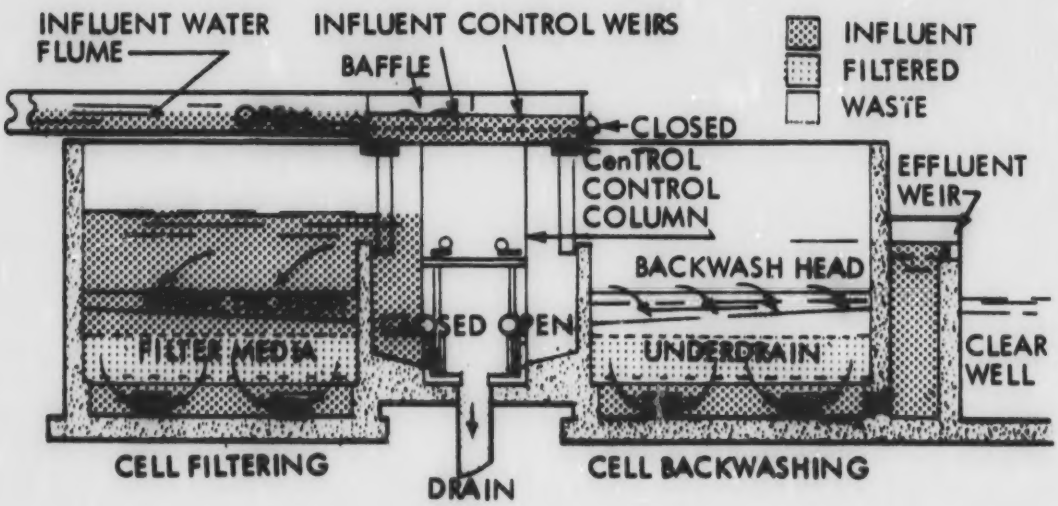


Figure 6b. Influent Flow Splitting - Filtration rate, water level, and head loss during two filter runs.



FILTERING CYCLE



BACKWASH CYCLE

Fig. 7. Operating Cycles of Influent-control, Constant Rate Filter.

In true declining rate filtration, the total pressure drop (head loss) across the filter media, underdrains and piping is held constant. This would be the case in Fig. 4 if the effluent rate control valve were adjusted at the beginning of the run and then the valve position were not changed throughout a filter run. In such operation, the rate of filtration is a maximum at the beginning of the run and declines as the run progresses. Declining rate filtration produces better water quality than constant rate filtration as evidenced by Fig. 8a and 8b. The advantages of declining rate filtration have also been reported by Hudson (6) and Tuepker (7).

True declining rate filtration is difficult to apply in actual water plants using gravity filters. Since each filter is declining independently, the sum of their several rates may not add up to the total plant input. Thus, the water level in the filter boxes may rise or fall and some method of control would be needed. However, a modification of true declining rate filtration is possible which has been named variable declining rate filtration (5).

Variable Declining Rate Operation

Variable declining rate operation is similar to influent flow splitting. Variable declining rate operation achieves all of the influent flow splitting advantages and some additional ones without any of the disadvantages. It can be adopted with little or no expense by hundreds of existing treatment plants whose controllers are not functioning properly or are causing excessive maintenance problems. Some plants have been designed for this method of operation. Others have been forced to do so due to inoperative rate controllers (6). However, despite the merits of this method, it has not received enough explanation or attention.

Fig. 9a illustrates the desirable arrangement for new plants designed for variable declining rate operation. Great similarity exists between Fig. 6a and 9a, the principal differences being the location and type of influent arrangement and the provision of less available head loss.

The method of operation is similar to that described for Fig. 6a with the following exceptions. Fig. 9b illustrates the typical water level variation and head loss variation observed with this mode of operation. The filter influent enters below the wash trough level of the filters. When the water level in the filters is <u>below</u> the level of the wash trough, the installation operates as an influent flow-splitting constant rate plant. When the water level is above the level of the wash trough, the installation operates as a variable, declining rate plant. In general, the only time the filter water level will be below the wash trough level will be when all filters are backwashed in rapid sequence or after the total plant has been shut down with no influent so the water level drops below the wash trough. In most cases, the clean filter head loss through the piping and underdrains will range from 3 to 4 ft. and keep the actual low water level above the wash trough. The water level is essentially the same in all operating filters at all times. This is achieved by providing a relatively large influent header (pipe or channel) to serve all the filters and a relatively large influent valve or gate to each individual filter. Thus, head losses along the header or through the influent valve are small and do not restrict the flow to each filter. The header and influent valve will be able to deliver whatever flow each individual filter is taking at the moment.

All of the filters in the bank of filters served by the common influent header get dirty; thus the flow through the dirtiest filters tends to decrease

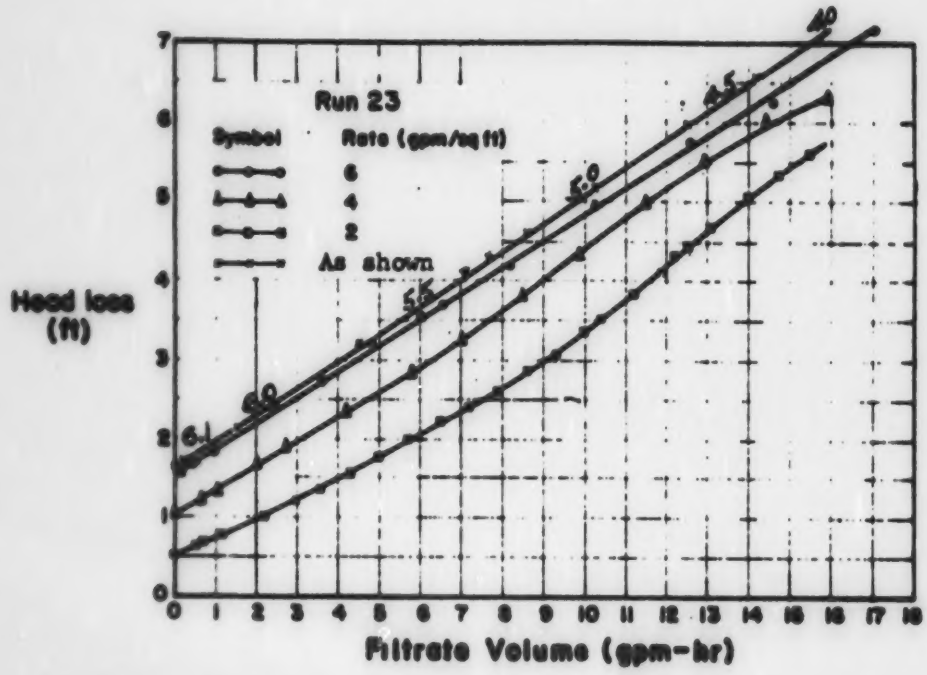
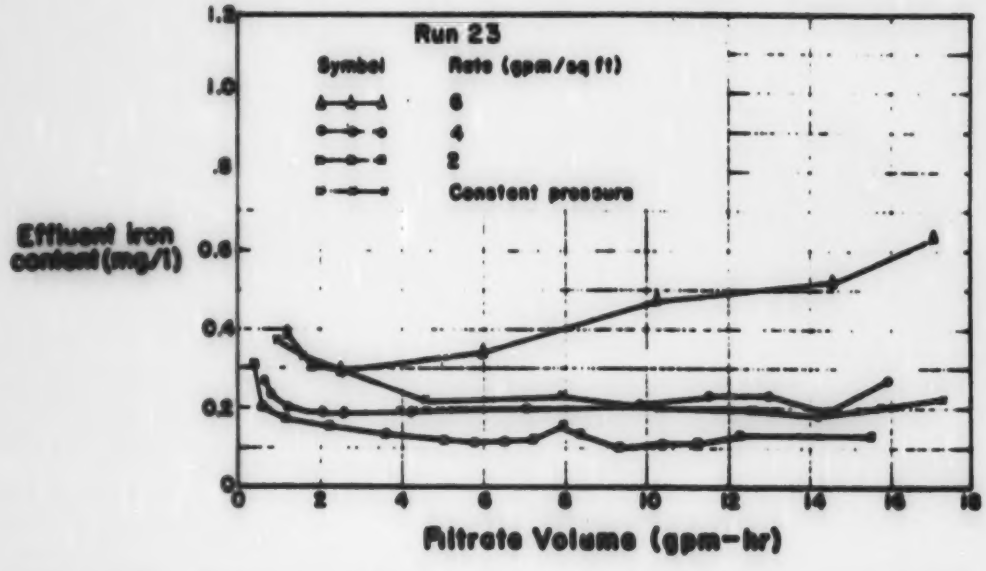


Figure 8a. Head loss comparing various constant rates and one declining rate.



Fibure 8b. Effluent quality comparing various constant rates and one declining rate.

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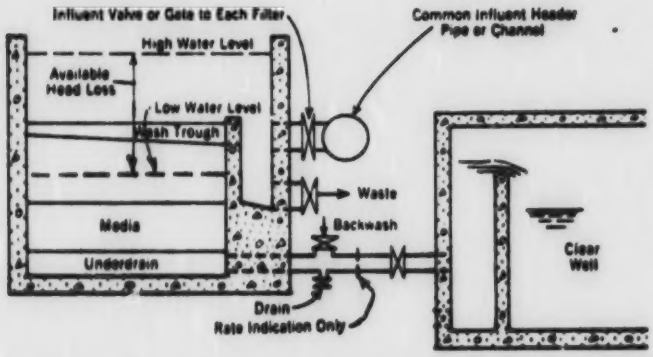

Figure 9a. Variable Declining Rate Filtration - Typical filter and clear well arrangement.

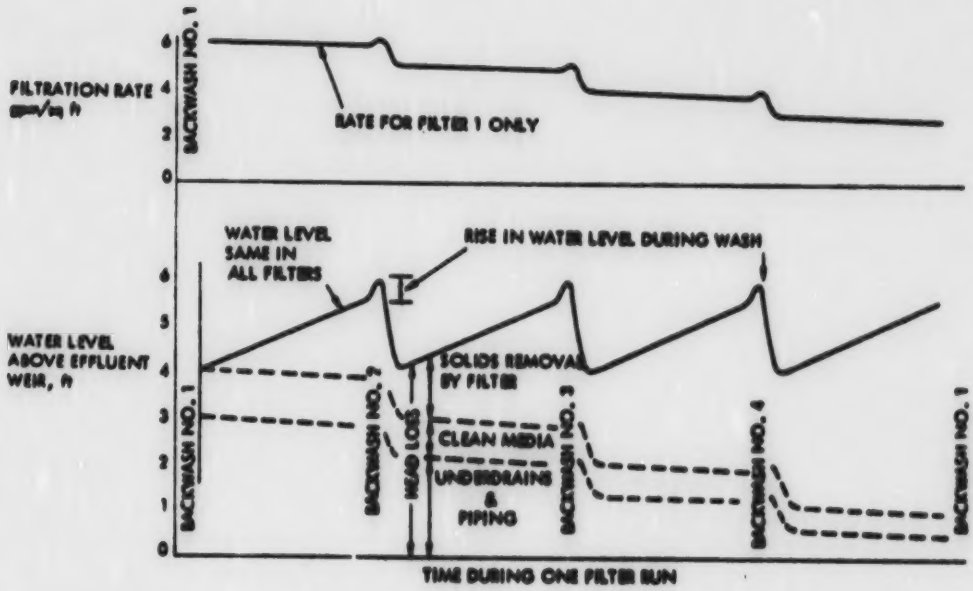

Figure 9b. Variable Declining Rate Filtration - Filtration rate, head loss, and water level during one filter run in a plant having four filters.

the most rapidly. This causes the flow to redistribute itself automatically so the cleaner filters pick up the capacity lost by the dirtier filters. The water level rises slightly in the redistribution of flow to provide the additional head needed by the cleaner filters to pick up the decreased flow of the dirtier filters. The cleanest filter accepts the greatest flow increase in this redistribution. As the water level rises, it partially offsets the decreased flow through the dirtier filters; as a result, the flow rate does not decrease as much or as rapidly as expected.

This method of operation causes a gradually declining rate towards the end of a filter run. Filter effluent quality is adversely affected by abrupt increases in the rate of flow -- here the rate increases occur in the cleaner filters where they have the least effect on filter effluent quality (8). Rate changes throughout the day due to changes in total plant flow, both upward and d[illegible]d, in all of the filters, dirty or clean, occur gradually and smoothly without any automatic control equipment.

The advantages of declining rate operation over constant operation are (5, 6, 7):

- Significantly better filter effluent quality than that obtained with constant-rate filter operation, and
- Less available head loss needed compared with that required for constant-rate operation.

The available head (or filter box depth) needed in operating a declining-rate filter is less than that needed to operate a constant-rate filter for three reasons:

- The flow rate through the filter decreases toward the end of the filter run. The head loss in the underdrain and effluent piping system therefore decreases and becomes available to sustain the run for a longer period than would be possible under constant-rate operation with the same available head. The recovery of available head due to decreased flow at the end of the filter run is a function of the square of the flow rate. (If the head loss through the underdrain and effluent system were 4 ft. at 4 gpm/sq.ft., it would decrease to 2.2 at 3 gpm/sq.ft. [$4 \times (3/4)2 = 2.2$] and provide an additional 1.8 ft. of head for continuation of the run on that filter).

- As the dirty filters lose capacity, the load is redistributed among all of the filters. The resulting rate increase in each filter is very small and the water level on all filters rises very slowly until it approaches the maximum permissible water level. The filter with the longest time in service in backwashed, and when put back in service, that filter operates at the highest flow rate thereby causing the water level on all filters to decline to a new equilibrium level.

- The influent enters below low water level thereby saving the head on the weir and the free fall allowance needed where influent-flow splitting is used.

For the above reasons, declining rate filters are considered to be most applicable to the type of filter operation encountered in wastewater filtration.

Filter Media

Granular filter media commonly used in water and wastewater filtration include silica sand, garnet sand, and anthracite coal. These media can be purchased in several effective sizes (that size that 10% by weight is smaller and 90% by weight is larger than) and several uniformity coefficients (the ratio of the size such that 60% by weight is smaller and the effective size or 60% size/10% size). The media have specific gravities approximately as follows:

Anthracite coal	1.4-1.6
Silica sand	2.65
Garnet sand	4.0-4.2

Media of these specific gravities are used so that after backwashing the media will be arranged with the coarse, lighter coal on top, the finer, heavier garnet sand on the bottom and the silica sand in between the two. The actual distribution of media after backwashing will depend, of course, on whether single, dual, or multi-media are used and the relative particle size distributions, and specific gravities of each media. Table 1 lists a number of suppliers of each media type.

The literature review (Appendix A) indicated that surface clogging of filters in wastewater filtration results in a rapid rate of head loss build-up and short filter runs. Surface clogging can be reduced by:

- using a narrowed range of media size, preferably a uni-sized media,
- using dual or multimedia filters, or
- using a coarse to fine flow direction (upflow) in a graded media filter.

The range in media size is quite important, a filter with a small range of sand size (low u.c.) provides longer run lengths (9). How narrow a range

Table 1

List of Media Supplies *

Anthracite Coal

1. Carbonite Filter Corporation
   Box #1
   Delano, Pennsylvania 18220

2. Shamokin Filler Company, Inc.
   Box 272
   Shamokin, Pennsylvania 17872

3. Reading Anthracite Coal Co.
   200 Mahantango Street
   P.O. Drawer F
   Pottsville, Pennsylvania 17901

4. Glenn Alden Fuel Sales
   P.O. Box 568
   Wilkes-Barre, Pennsylvania 18703

Silica Sand and Gravel

1. Eau Clair Sand and Gravel Company
   104 Gibson Street
   Eau Claire, Wisconsin 54701

2. Northern Gravel Company
   Box 307
   Muscatine, Iowa 52761

3. Jesse S. Morie & Sons, Inc.
   Mauricetown, New Jersey 08329

4. The Parry Company
   219 Church Street
   Chillicothe, Ohio 45601

5. Monterey Sand Company, Inc.
   Box 928
   Monterey, California

6. Cosby-Carmichael, Inc.
   Box 597
   Selma, Alabama 36701

7. William Bird Sales Corporation
   Box 5442
   Columbia, South Carolina

8. Southern Products & Silica Company
   Box 38
   Lilesville, North Carolina 28091

9. George F. Pettinos, Inc.
   235 Bala Avenue
   Bala-Cynwyd, Pennsylvania 19004

10. New Jersey Pulverizing Company
    205 West 34th Street
    New York 10001

Garnet Sand

1. Idaho Garnet Abrasive Company
   Box 1080
   Kellogg, Idaho 83837

*This list is not a complete list of suppliers and is included for general information only.

is desired has not yet been determined. The best u.c. range is influenced to some extent, by the pretreatment given the wastewater and the solids level to be removed.

Uni-sized media filters have been operated and investigated primarily on a research basis (10-19). The main reason for using a uniform grain size originally in the research was the simplicity of predicting the filter performance. A systematic study of iron removal comparing the filter performance of graded media and uni-sized media was conducted by Kim in Alaska (38). He claimed that a properly designed uni-sized media could increase run length more than ten times over that of graded sand filters of the same effective size. Apparently, uni-sized sand reduced considerably or eliminated the surface clogging problem.

Use of uni-sized media will increase the cost of media. However, the portion of filtration cost contributed by the filter media is only a small portion of the total cost of a filter plant. Therefore, the added cost of a uni-sized media is not a valid argument against its use.

A second approach to reduce surface clogging is through the use of dual or multi-media filters. Conley (21) advocated the use of a dual media filter composed of coarse anthracite on top of finer sand to attain filtration in the direction of diminishing grain size. Ives (22) reported some results of triple-media filtration, in which a third material, garnet sand, was installed beneath the silica sand. Mohanka (23) even tested a five-layer filter using polystyrene, anthracite, sand, garnet, and magnetite located from top to bottom in the order listed. Triple- and five-media filter showed, respectively, 33-67 percent and 37-60 percent of the headloss of a sand bed of equal depth. In all these modifications, the fine grain at the top of each medium cannot be avoided if graded materials

are used. Also, the use of a very heavy, small grain at the bottom requires extreme care in the composition of the supporting gravel, lest underdrain clogging occur. Thus, a <u>further</u> beneficial modification to dual media filters should be made by using as close to uni-sized material as possible in each layer.

It should be noted that a uniform size of medium is impossible to achieve in practice. However, a medium graded into the narrowest size range of available sieves should result in a nearly uniform sized medium. A more nearly uni-sized media can be obtained in two ways:

- By purchase of a media that passes one standard Tyler or U.S. Standard sieve and is retained on the next smaller sieve, or
- By purchase of a standard media of the desired effective size and by "scalping" the fines off the media as it is placed on the filter.

For example, a recent study (24) indicated that for the filtration of a trickling filter effluent, a recommended satisfactory filter media could consist of 15 inches of 1.84 mm uni-size anthrafilt on top of 15 inches of 0.55 mm uni-size sand. To prepare these media, Philterkol Special No.1* filter media (effective size 1.0-1.1 mm, u.c. max. 1.80) was sieved through a Gilman set of U.S. Standard sieves on a mechanical shaker for 5 minutes. The filter sand** (0.55-0.60 mm effective size, u.c. of 1.7) was sieved in a similar manner for 10 minutes.

The uni-sized sand or anthracite used was that media 100 percent of which passed one sieve number and was retained on the next sieve. For example, a 10/12 anthracite was one in which 100 percent of the material passed a U.S. Sieve No. 10 and 100 percent was retained on a U.S. No. 12.

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* Philterkol Special No. 1, Reading Anthracite Coal Company, Pottsville, Pennsylvania.

** Northern Gravel Co., Muscatine, Iowa.

A screen analysis of the commercially graded anthracite, from which a uni-sized 1.84 mm anthracite (passing U.S. Sieve No. 10 and retained on U.S. Sieve No. 12) was obtained, is shown in Table 2.

Table 3 shows the geometric mean sizes of uni-size filter media that can be obtained in this manner. Table 2 indicates that only 22.9 percent of a shipment of the Philterkol Special No. 1 media will be useable as a uni-size 1.84 mm media (pass U.S. No. 10 Sieve and retained on U.S. No. 12 Sieve). A total of 45.6 percent would be useable between the U.S. No. 10 and the U.S. No. 14 Sieves if a slightly finer media is desired and a total of 42.3 percent would be useable between the U.S. No. 8 and the U.S. No. 12 Sieves. Uni-sized media are desired principally to get rid of the very fine material which will collect on the filter surface and cause removal of suspended solids with head loss results to be described in a later section.

A better wastewater filter can also be obtained by ordering the desired effective size of filter media with as small a uniformity coefficient (1.5-1.7) as is readily commercially available. Then, as each media is placed in the filter, it should be backwashed several times to grade the media so that the fines are at the top. Then, a predetermined depth of media can be skimmed (or scalped) off the surface and discarded. Scalping of the sand would be followed by the placement of the anthracite media and the backwashing of the filter to again grade the media so the fines are on top. In the case of Philterkol Special No. 1 (Table 2), the top media would have an average size of less than 0.297 mm and would undoubtedly cause a surface removal of suspended solids. After backwashing and draining, the skimming off of the top 9 percent of the media (about 2 inches out of 20 inches for example) would reduce the coal size in the surface to 1.00 mm

Table 2. - Screen analysis of Philterkol Special No. 1*

Openings		Tyler Mesh	U.S. No.	Per Cent Weight Retained	Per Cent Cumulative Weights
Inches	Milli-meters				
.132	3.36	6	6		
.0937	2.33	8	8	3.0	3.0
.0787	2.00	9	10	19.4	22.4
.0661	1.68	10	12	22.9	45.3
.0555	1.41	12	14	22.7	68.0
.0469	1.19	14	16	14.0	82.0
.0394	1.00	16	18	9.8	91.8
.0331	0.841	20	20	5.2	97.0
.0278	0.707	24	25	2.2	99.2
.0234	0.595	28	30	0.2	99.4
.0165	0.420	35	40	0.2	99.6
.0117	0.297	48	50	0.2	99.8
		-48	-50	0.2	100.0

* E.S. 1.00 to 1.10 mm, U.C. max 1.80.

Table 3. Size range of uni-sized sand or anthracite

Geometric mean size mm	Sieve opening		U.S. sieve No.	
	Passing mm	Retained mm	Passing	Retained
0.46	0.50	0.42	35	40
0.55	0.60	0.50	30	35
0.65	0.71	0.60	25	30
0.77	0.84	0.71	20	25
0.92	1.00	0.84	18	20
1.09	1.19	1.00	16	18
1.30	1.41	1.19	14	16
1.54	1.68	1.41	12	14
1.84	2.00	1.68	10	12

(Table 2). Skimming off 18 percent of the media (about 4 inches out of 20 inches) would reduce the surface coal size to about 1.19 mm. This practice should be followed in placing filter media for wastewater filtration unless a uni-size media becomes more readily commercially available.

Backwashing requirements

Filters must be washed to restore their filtering capacity when the filtrate quality becomes unacceptable or when the head loss through the filter reaches the maximum head available (Fig. 1). In many cases, filters are backwashed on a regular time cycle based on the operating experience with these two criteria at the installation. Filters are washed by reversing the flow of water (usually filtered water with downflow filters and raw, unfiltered water with upflow filters) at a rate sufficient to lift the media grains into suspension. The rising water thus expands the bed allowing the suspended solids collected in it to be washed up through the expanded bed and to waste through the wash water gutters. Recent studies have shown that the abrasion and collision of particles in the expanded bed is limited (25, 26). This means that the chief mechanism involved in particle cleaning by water wash along is due to the hydrodynamic shear and turbulence generated by flow of water around the expanded particles. Since both hydrodynamic shear and turbulence forces are at a maximum when the void ratio in the expanded bed reaches a valve of 0.70, we may expect best filter cleaning when the expansion produces a void ratio of 0.70 near the surface of the media where most of the suspended solids are collected. A recent paper has presented several models which can be used to predict the expansion of sand filter media within close limits. The models, however, cannot yet predict the expansion of coal with equal precision (26).

In order to expand the surface layer of a graded sand (effective size of 0.45 mm and u.c. of 1.47) to a porosity of 0.71 would require a 49 percent

expansion of the bed (27). In a dual media filter, the coal layer would normally have a void ratio of about 0.50. To expand this to a void ratio of 0.70 would require a 20-25 percent expansion of the media. Thus, Amirtharajah predicts (and suggests studies to prove) that optimum water wash of dual media filters would occur with an expansion of 20-25 percent. He also concludes that water wash alone is not sufficient to provide the desired degree of media cleaning since water alone is not effective in producing abrasion of the media to release the attached suspended solids. Accordingly, it may be concluded that all wastewater filters should be provided with a preliminary air wash to produce the particle to particle contact for effective media cleaning.

U.S. practice has tended to use a high velocity backwash (commonly 15 to 19 gpm/sq.ft. for 3 to 10 minutes) for washing sand filters. This results in about 25-30 percent expansion of the sands currently in use, depending on water temperature, media gradation, and specific gravity. Early work by Hulbert and Herring found that a high velocity wash sufficient to get 50 percent sand expansion gave the best washing results (28) thus confirming the current work (27). However, our recent movement to use of coarser media resulted in excessively high wash rates at that degree of expansion. As a result, current practice now tends to use of lower wash rates preceded by an air wash.

At present, the desired backwashing sequence for a dual media filter for wastewater applications would include the following:

An air wash of 3 to 5 CFM/sq.ft. for a period of 3-4 minutes.

A water wash sufficient to produce about a 25 percent expansion of the dual media for a period of 5-10 minutes at the highest water temperature to be encountered in normal practice.

Since it is not now possible to predict coal expansion with an accuracy better than 15 percent, the backwashing rate should be established

on the basis of a laboratory test. It is relatively simple to set up a 4-6 inch laboratory column containing the single or dual media to be used and to determine the media expansion at various backwash rates at the critical wastewater temperature. Figure 10 shows the results of such a study to determine the backwash rate needed to expand dual filter media consisting of uni-size 1.84 mm coal on top of 0.77, 0.65, and 0.55 mm uni-size sand.

The time required for backwashing a filter is defined as the total period of time when a filter is down for backwashing, including the time for (1) draining down the water above the filter media to a level about 1 foot below the wash trough; (2) air scouring; (3) water washing; and (4) leeway. It is common practice to estimate backwash time to be a half hour. However, the backwash time depends on the filtration rate used as illustrated in the following example:

In this example, the terminal headloss is assumed to be 10 ft. water, which is the water column above the surface of the filter media, and the wash trough level is assumed to be 4 ft. above the surface of the filter media, as shown in Fig. 11. When the headloss reaches the pre-determined 10 ft. water level, the filtration cycle is terminated and the backwashing cycle is started. Before initiating the air scouring, the residual water above the filter media must be filtered down to a level at least one foot below the wash trough in order not to waste this water and to provide room for a water level rise during air wash. An air wash of 3-5 CFM/sq.ft. will result in a water level rise of about 6-9 inches. The time period required depends on the filtration rate used. If a uniform decreasing of the draining rate from the terminal headloss water level to the surface of the filter media is assumed, an average draining rate can be estimated and the

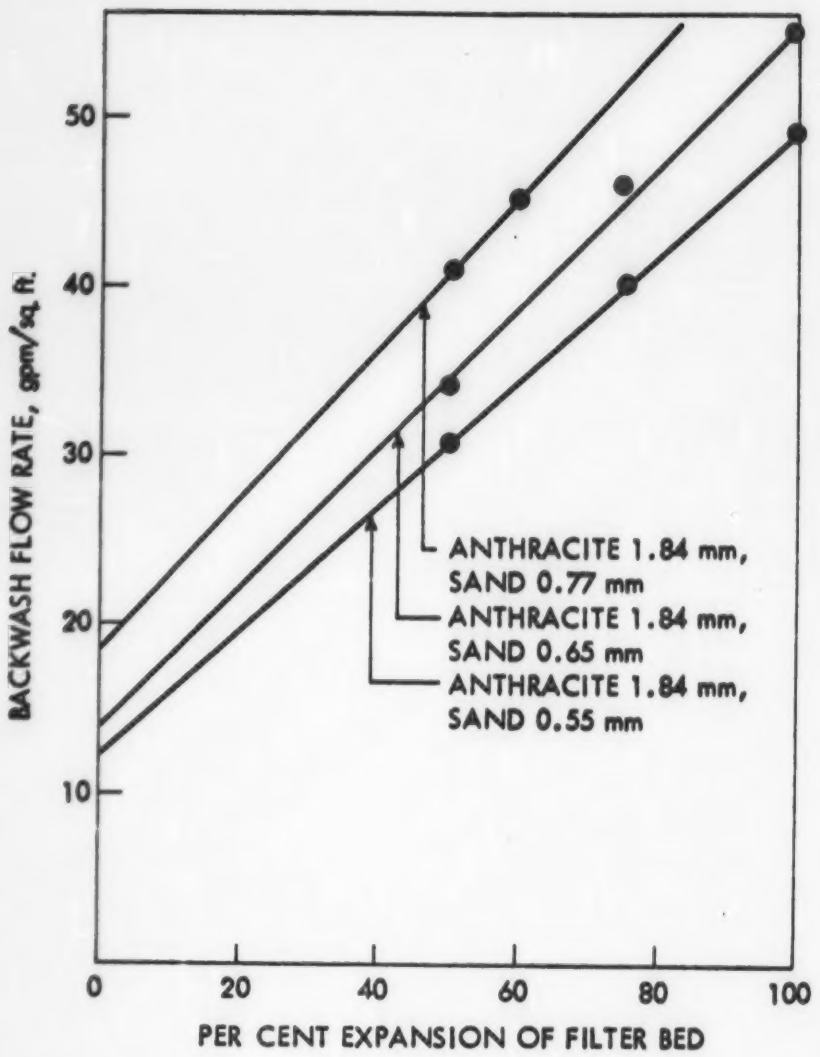


Fig. 10. Media bed per cent expansion vs. backwash flow rate at various media size combinations (water temperature 50-60°F)

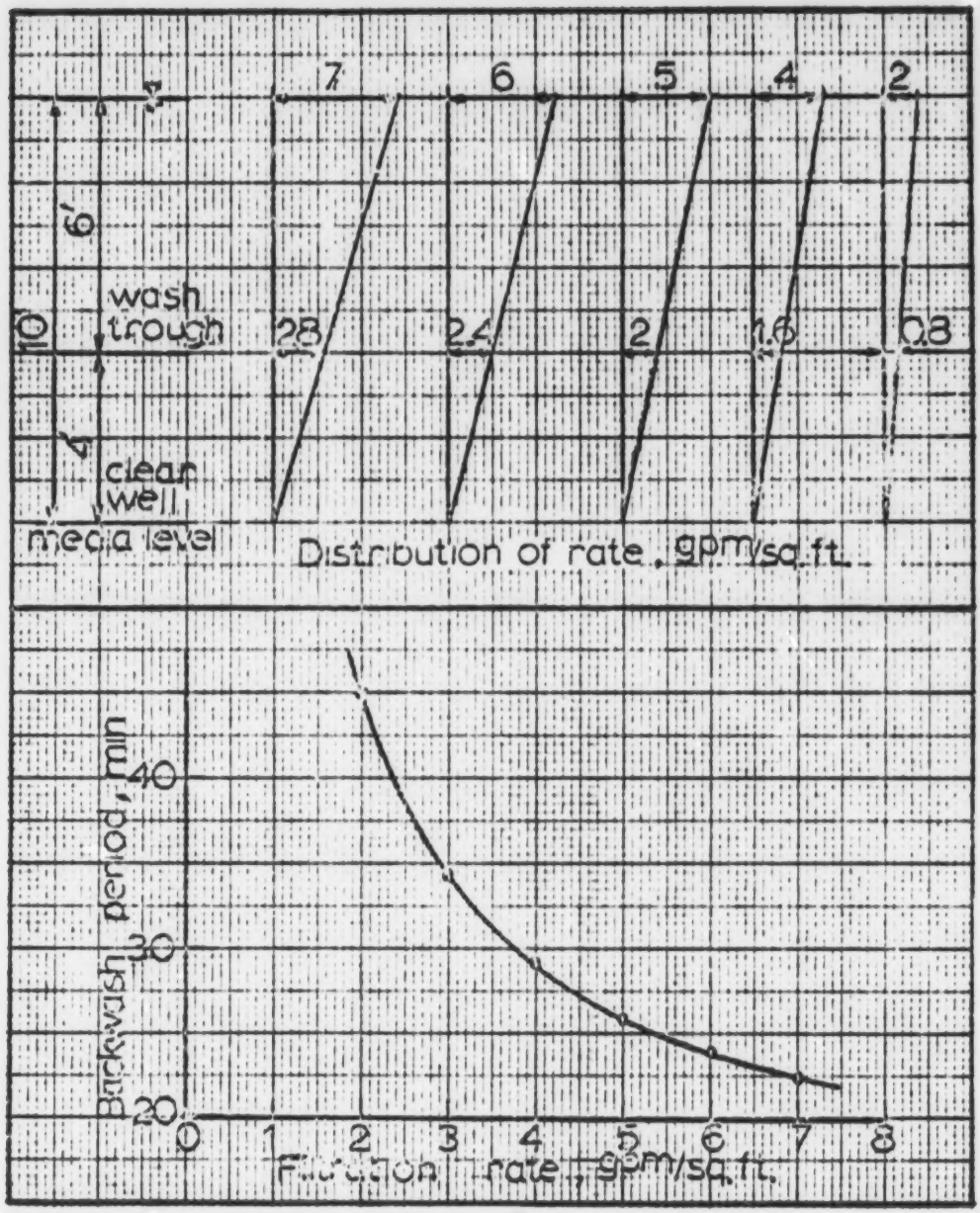


Fig. 11. Theoretical backwashing period vs. filtration rate

time required for filtering down to the level of the wash trough is obtained as follows:

At a filtration rate of 7 gpm/sq.ft.:

Average draining rate =

$$\frac{7(\text{gpm/sq.ft.}) + 2.8(\text{gpm/sq.ft.})}{2} = 4.9 \text{ gpm/sq.ft.}$$

Time required for draining =

$$\frac{6(\text{cu.ft.}) \times 7.48(\text{gal/cu.ft.})}{4.9 (\text{gpm/sq.ft.})} = 9.15 \text{ min.}$$

Therefore, total down time will be

drain time to wash trough	9.15 min.
air scouring	3.00 min.
water washing	5.00 min.
leeway	5.00 min. (time to drain 1-ft. below wash trough plus some leeway)
	22.15 min.

A similar analysis can be made for filtration at rates other than 7 gpm/sq.ft. The results, shown in the lower portion of Fig. 11, indicate that the period required for filter backwashing increases exponentially as the filtration rate decreases. The typical periods of down time would be:

Filtration rate gpm/sq.ft.	Average drain rate gpm/sq. ft.	Down time min.
2	1.4	45.0
3	2.1	34.4
4	2.8	29.0
5	3.5	25.8
6	4.2	23.7
7	4.9	22.15

The relationship between net water production and filtration rate at various run lengths is shown in Figure 2 based on this calculated time required for backwash.

If shorter filter down times for backwashing are desired, the time for draining the filter down to the wash trough level can be reduced to less than one minute by dumping the water through the wash trough back to the influent side of the plant. Time would still be required for dropping the water level one foot below the trough by filtering, however, to provide space for water rise during the air wash. Air wash and water wash should not be used simultaneously unless the air wash is shut off when the water level is at least one foot below the wash trough. Even if the water were drained to the media surface, only a short air-water wash could be used:

Total head below wash trough ----------	4.0 ft.
Required leeway at air-wash shut-off --- 1.0 ft. -----	1.0 ft.
Required depth for water expansion during air wash - 0.75 ft. ----------	0.75 ft.
Head available for air-water wash ----------	2.25 ft.
Assumed backwash rate ----------	24.0 gpm/sq.ft. or rise of 3.2 ft/min.
Time for air-water wash ---------- $\frac{2.25 \text{ ft.}}{3.20 \text{ ft/min.}}$ = ----	0.7 min.
	(Impractical)

Terminal Head loss and Head loss Development

Modeling of the filtration process has not yet progressed to the point where it is possible to determine precisely what economic terminal head loss should be provided across a granular media filter. Huang and Baumann (2) recently found that the most economic terminal head loss for filtration of iron on uni-sized sand filters ranged between 8 and 11 feet (Fig. 12). Normal American practice would use a terminal head loss of 8 to 10 feet when using gravity filters. The available head should not be so high as to result

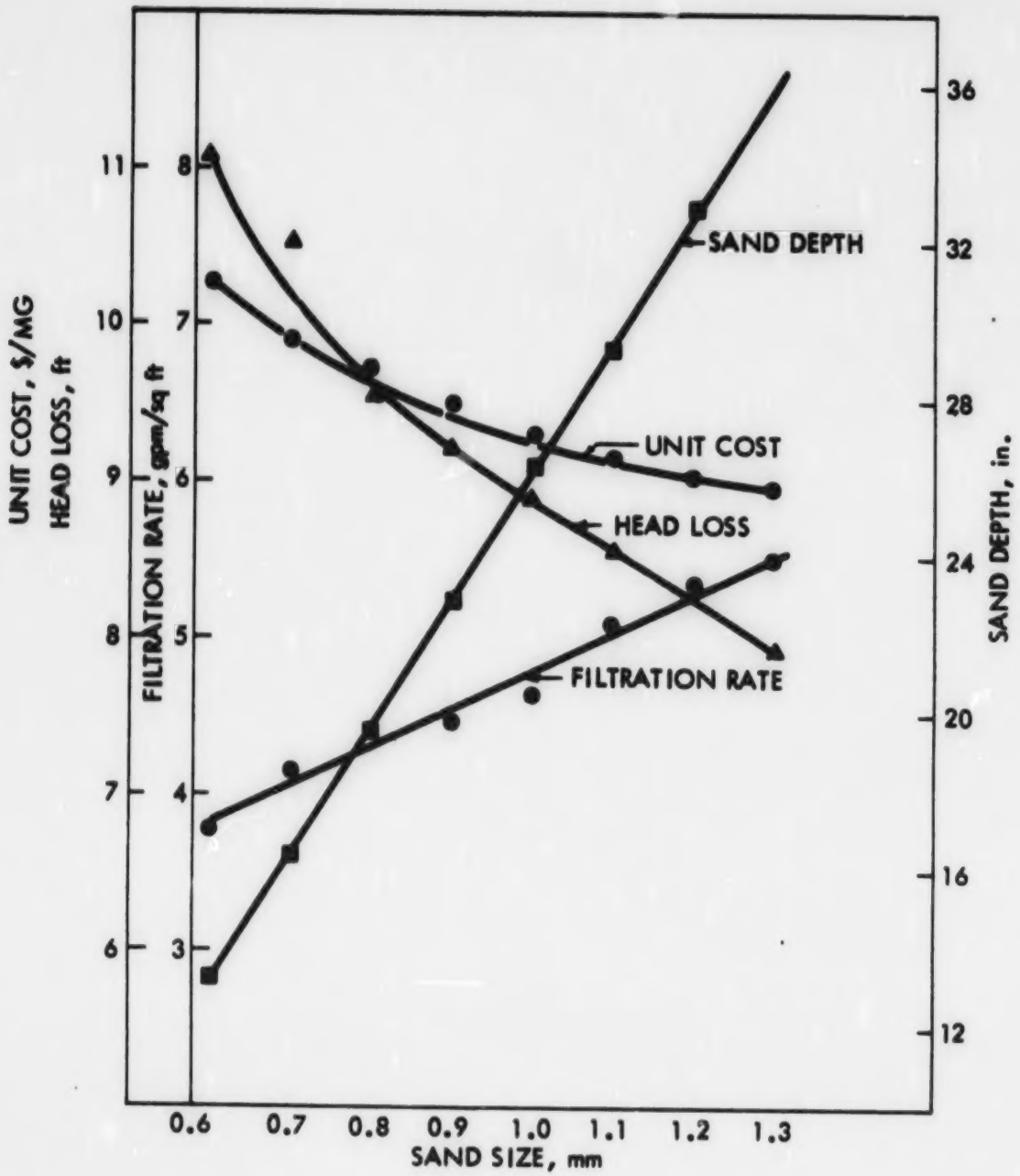


Fig. 12. Most Economic Filter Design Combinations for Removal of Iron on Uni-size Sand Filters.

in failure of the filtration process by solids breakthrough (Fig. 1). In general, this will mean provision of available head of 8-12 feet, with this author preferring about 10 ft.

A recent study indicated that pressure drops of as much as 30 ft. of water could be used in plain filtration of trickling filter effluents through dual media filters without solids breakthrough. However, economic considerations indicate a maximum practical head limit of 10-12 ft.

Figure 13 shows several examples of head loss development during solids separation by filtration. Relatively speaking, granular filters remove suspended solids in one of the following ways:

- By removal of the suspended solids at the surface by the finer media at the top of the filter. This forms a relatively thin layer of deposited solids at the surface.
- By depth removal of the suspended solids within the voids of the porous media. The better the distribution of the solids throughout the depth of the filter media, the better the use of the head available.
- By a combination of surface removal and depth removal.

Figure 13 includes a schematic section through each filter cake and a typical head loss vs. volume of filtrate relationship observed using that type of filter.

With surface removal of suspended solids, the filter medium can be a piece of filter paper or membrane in the laboratory, the "Schmutzdecke" skin in a slow sand filter, a synthetic fabric or a fabric precoated with a filter aid such as diatomite or perlite or powdered activated carbon or mixtures of these materials. The filter medium can even be the surface of a bed of fine-grained sand or coal. Early in the filter run, with surface removal of suspended solids the filtration resistance of the filter depends on the filter porosity, thickness of the medium, and so forth. Once a surface layer of the suspended solids forms on the filter medium, however, the

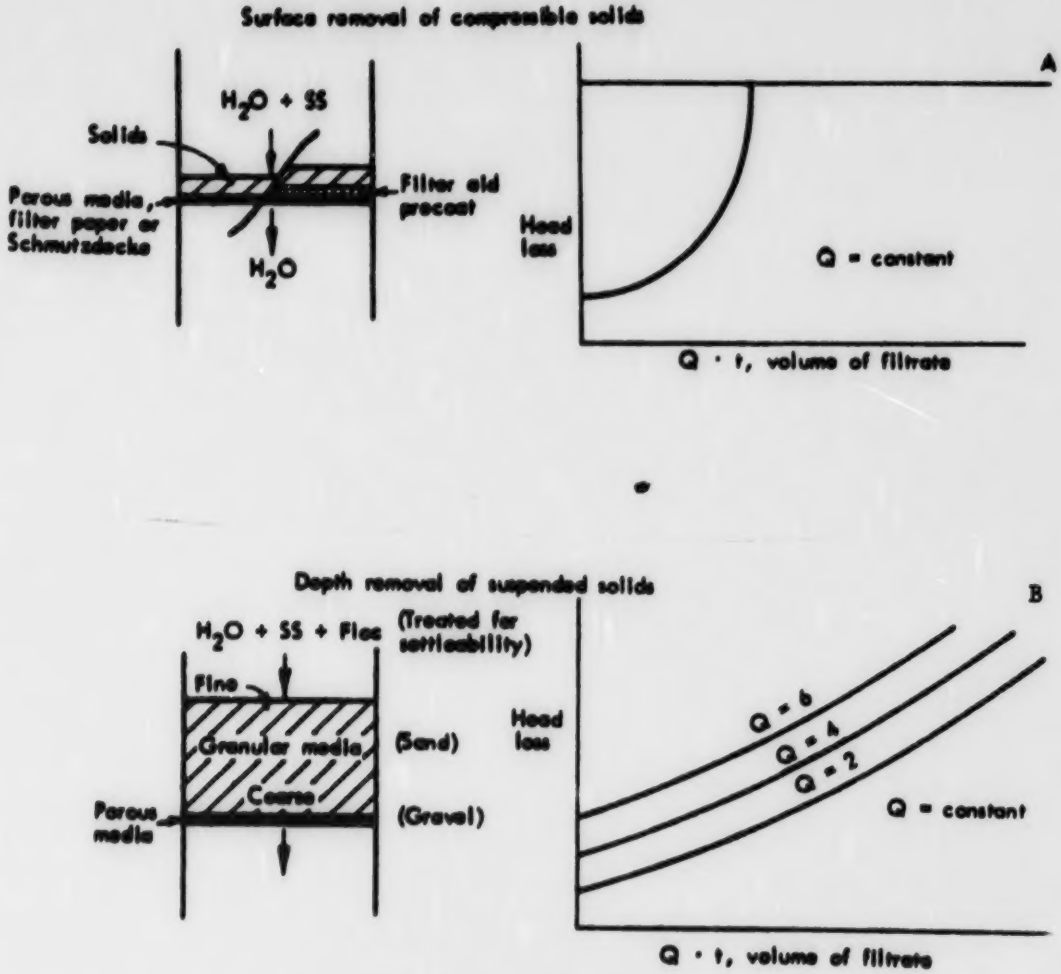
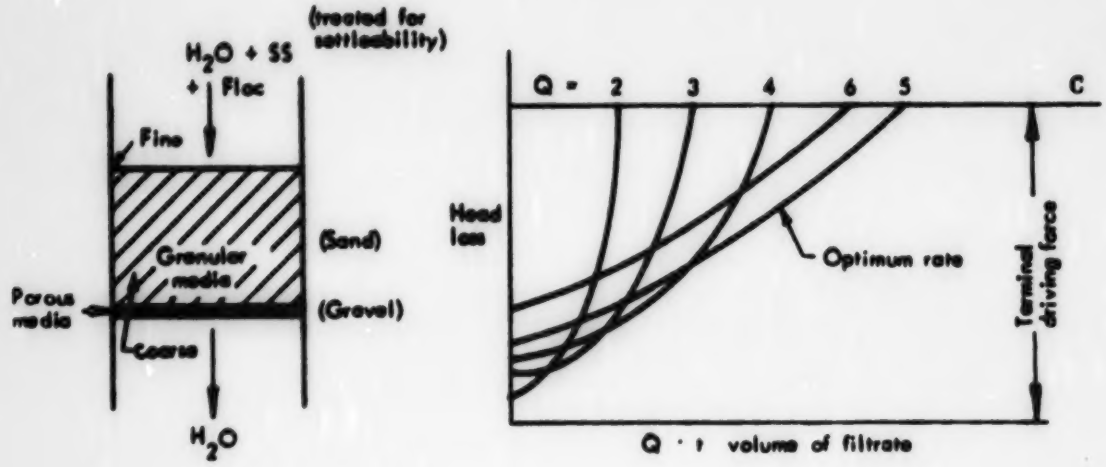


Fig. 13. Head Loss Development During Filtration.

Depth removal of suspended solids with surface cake



Depth removal with reverse graded granular media

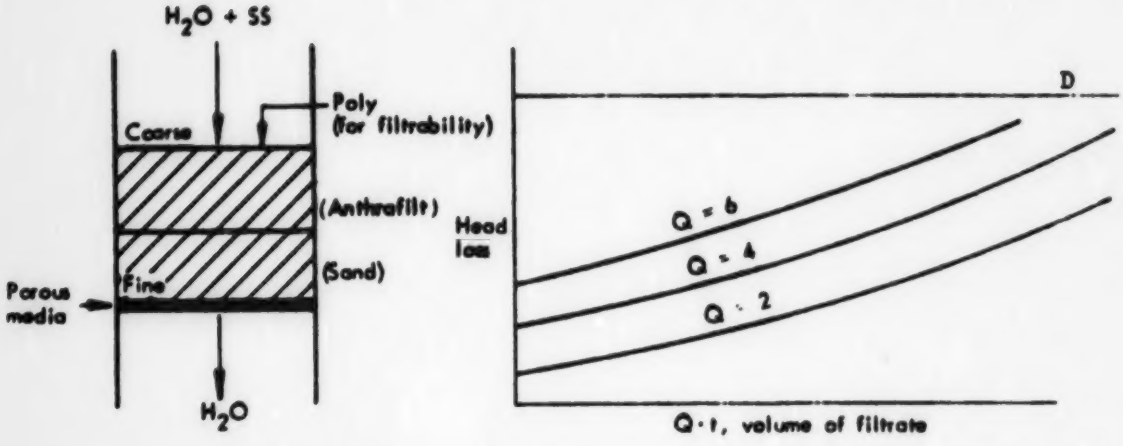


Fig. 13. Head Loss Development During Filtration.

removal of additional suspended solids is accomplished by filtration through the solids already removed. Thereafter, the filter resistance changes are due to changes in the thickness and resistance of the surface layer of solids forming on the filter. Once the surface layer of solids has formed, the relative efficiency and operating characteristics of all such filters are the same and can be predicted based on a knowledge of the solids resistance characteristics and the filter area, water temperature, and available driving force. With this type of filter, it is usually desirable to keep the driving force relatively low to prevent cake compression and to maintain as porous a cake as possible. Adequate filter capacity can be achieved by keeping flow rates very small by using as large a filter area as possible to meet the demand rate for filtration.

As solids build up on the filter surface, the thickness of the cake and the pressure across the cake both increase. As a result, the resistance of the cake increases due to compression of the cake. The head loss generated across the filter will increase slowly at first (Fig. 13A) and then will increase more rapidly with time or volume of filtrate. With this type of surface removal, increasing the pressure drop does not significantly increase filter capacity.

Figure 13B shows the kind of suspended solids removal that is obtained with depth in a granular filter medium. These filters operate at relatively high filtration rates (2 to 10 gpm/ft.^{2}) on water that has been given a chemical pretreatment to improve settling (and in some places, for filtrability). If the residual solids are so fine that there is no surface removal by the fine medium on the top of a granular medium filter, solids removal depends on transport of solids from the bulk of the fluid (the transport step) to the surface of the medium (the attachment step). The

efficiency of the filter is determined by the effectiveness of the transport and attachment steps. With this type of filter, interruptions in the filtration cycle frequently release large quantities of once-removed solids back into the filter effluent (8).

With the absence of surface removal, increasing the flow rate will both increase the initial loss through the clean filter and distribute the solids removed through a larger depth of the granular medium. As a result, the head loss vs. volume of filtrate relationship will develop as shown in Figure 13B. Higher filtration rates will provide nearly the same (but decreased) total volume of filtrate per run, and the length of run will be much reduced. For a given suspension and a uniform sand of size d, if the depth of filter bed and time of filtration are fixed:

Q is proportional to $(1/d)^2$ for comparable quality, and

Q is proportional to $d^{1.3}$ for comparable head loss.

It is evident that the effect of media size on quality and head loss work in opposition to each other.

When the solids are partly removed on the surface and partly in the depth of the filter, surface removal will predominate at low filtration rates and the head loss generated is characteristic of surface removal head losses (Figure 13A). With higher rates, the solids are carried deeper into the granular medium and more filtrate is produced before the surface cake forms (Q = 3 and 4, Fig. 13C). The rate may become high enough to prevent surface cake formation and the head loss will then be controlled only by depth filtration. In Figure 13C, therefore, a flow rate of 5 would be the optimum rate since it produces the most filtrate per run. No surface cake forms with filtration rates of 5 or higher.

In recent years, upsidedown or reverse-graded filters using dual or

multi-media have been developed using the coarse material on the raw water side of the filter and the fine material on the filtered water side of the filter. Such filters distribute the solids much more uniformly throughout the depth of the filter so much of the solids will be removed in the coarse material. These filters may operate on water that has not been given pretreatment if the suspended solids in the water to be removed are less than about 100 or with pretreatment to improve settling and/or filtrability if phosphorous is to be precipitated. The head loss in these filters is generated at a much lower rate than that of a conventional filter. Filtration through a reverse-graded medium will provide a length of filter run that is 5 to 20 times greater than would be obtained with the same filter operated in the conventional manner. Figure 13D shows the typical head loss vs. volume of flow relationship for a reverse-graded filter. Since the water is being filtered through a medium of increasing fineness this type of filter is less subject to passage of solids due to filtration interruptions. Some breakthrough of solids will occur, but not to the same extent as with a conventional filter.

Filtration Rate

Filtration rates of 0.6 to 20 or more gpm/sq.ft. have been used in wastewater filtration (See Appendix A). Today, most commonly used wastewater filtration rates range between 3 and 6 gpm/sq.ft. Increased filtration rates both shorten filter runs and reduce filtrate quality (Fig. 8b). In some applications (Fig. 13c), a higher filtration rate may be necessary to overcome surface cake formation and the existence of an optimum filtration rate. The filtration rate must be:

- Kept low enough to produce the desired effluent quality desired consistent with the filter media adopted.

- Kept high enough to increase net water produced (and reduce filtration costs) consistent with good operating practice.

Figure 2 and Table 4 can be used to demonstrate how the effectiveness of filtration rate can be evaluated. Figure 14 shows a plot of run length obtained during filtration (at rates of 2, 4, and 6 gpm/sq.ft. to a 10 ft. terminal loss) of a trickling filter plant effluent at different suspended solids levels. Figure 15 shows the same data plotted in a more useful way. Using the data in Figure 15, let us consider what filtration rate might be used to remove the SS from the trickling filter effluent when the SS level is at a maximum of 40 mg/l. Run lengths from Figure 15 would be:

Rate gpm/sq.ft.	Run length, hrs.
2	20
4	12.5
6	8.5

At 2 gpm/sq.ft., the filter would produce a net of 2,690 gal of filtrate/sq.ft/day (Table 4). Even an infinite run length would increase net production to only 2880 gpd/sq.ft. At a filtration rate of 4 gpm/sq.ft., a 12.5 hour run would produce a net filtrate production between 5,260 (10 hr. run) and 5,500 (20 hr. run) gpd/sq.ft. and the use of the higher filtration rate would increase production by (5,300/2,690) about 1.98 times. Since run lengths could drop as low as 5 hours without serious loss of net production, this filtration would be more economical and still practical as compared to the 2 gpm/sq.ft. rate.

At a filtration rate of 6 gpm/sq.ft., a run length of about 8.5 hours would be obtained. Here, net production would be 7,440 gpd/sq.ft. at a run length of 5 hours and 8,000 gallons at a run length of 10 hours. Average net production (Table 4 or Figure 2) would be about 7,800 gallons or about 2.9 times the production at the 2 gpm/sq.ft. filtration rate. With normal variations, actual run lengths might be expected to vary between 5 and 10

Table 4. Net water productions determined by flow rates and run lengths

Flow rate gpm/sq.ft.	Run length hr.	Water prod. per run gal/sq.ft.	Backwash water* gal/sq.ft.	No. of runs** per day	Net water production gpd/sq.ft.
	∞				2,880
	50	6,000	100	0.475	2,800
	30	3,600	100	0.788	2,760
	20	2,400	100	1.170	2,690
2	10	1,200	100	2.285	2,510
	5	600	100	4.370	2,180
	3	360	100	6.85	1,780
	2	240	100	9.60	1,345
	1	120	100	16.00	320
	∞				5,760
	50	12,000	100	0.475	5,650
	30	7,200	100	0.788	5,570
	20	4,800	100	1.170	5,500
4	10	2,400	100	2.285	5,260
	5	1,200	100	4.370	4,810
	3	720	100	6.850	4,250
	2	480	100	9.60	3,640
	1	240	100	16.00	2,240

* 5 min. water backwash at rate of 20 gpm/sq. ft.
** 30 min. down time per run assumed.

Table 4. (continued)

Flow rate gpm/sq.ft.	Run length hr.	Water prod. per run gal/sq.ft.	Backwash water* gal/sq.ft.	No. of runs** per day	Net water production gpd/sq.ft.
	∞				8,640
	50	18,000	100	0.475	8,500
	30	10,800	100	0.788	8,440
	20	7,200	100	1.170	8,410
6	10	3,600	100	2.285	8,000
	5	1,800	100	4.370	7,440
	3	1,080	100	6.850	6,700
	2	720	100	9.60	5,950
	1	360	100	16.00	4,160

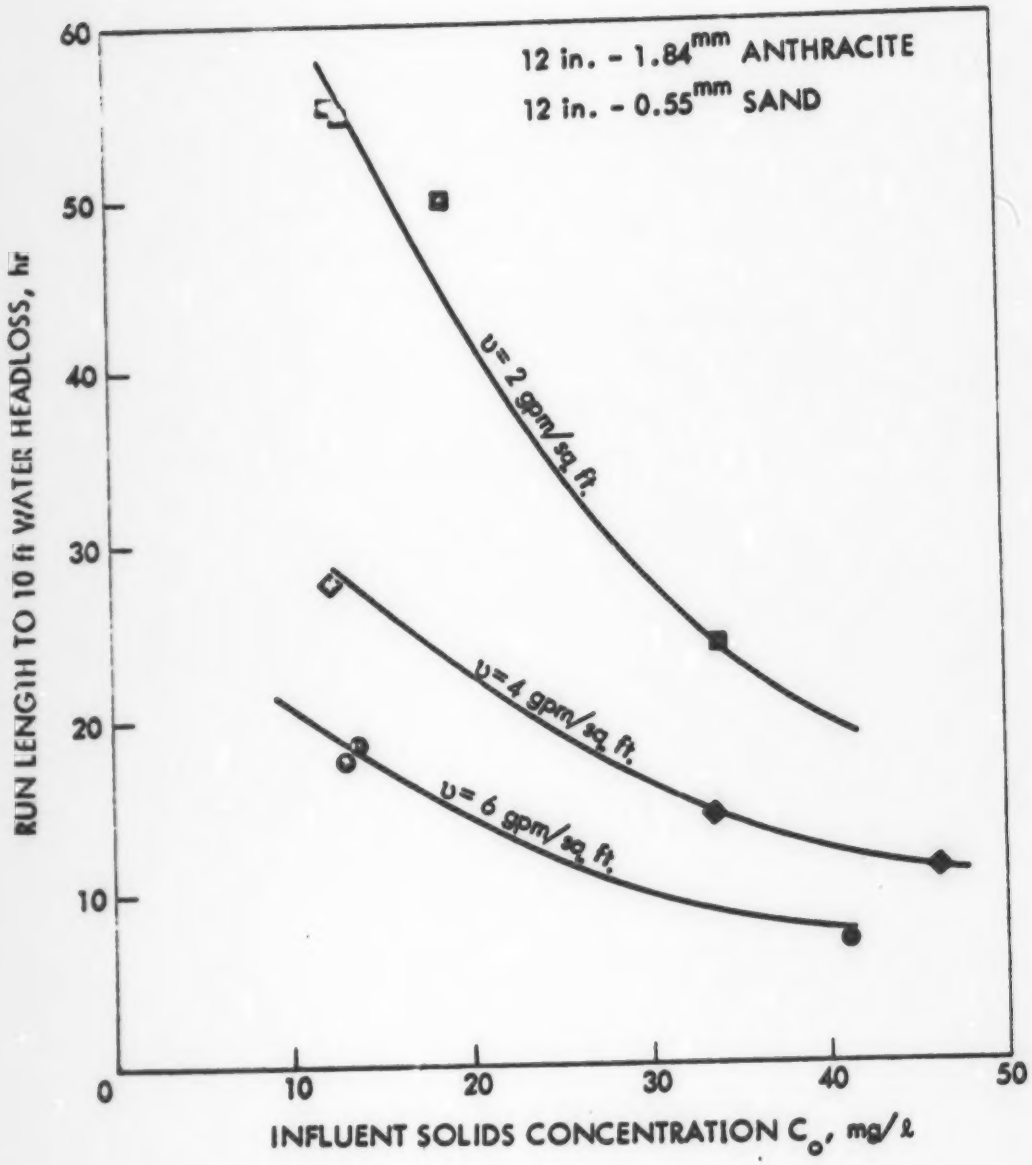


Fig. 14. Run length vs. influent SS concentration at various flow rates

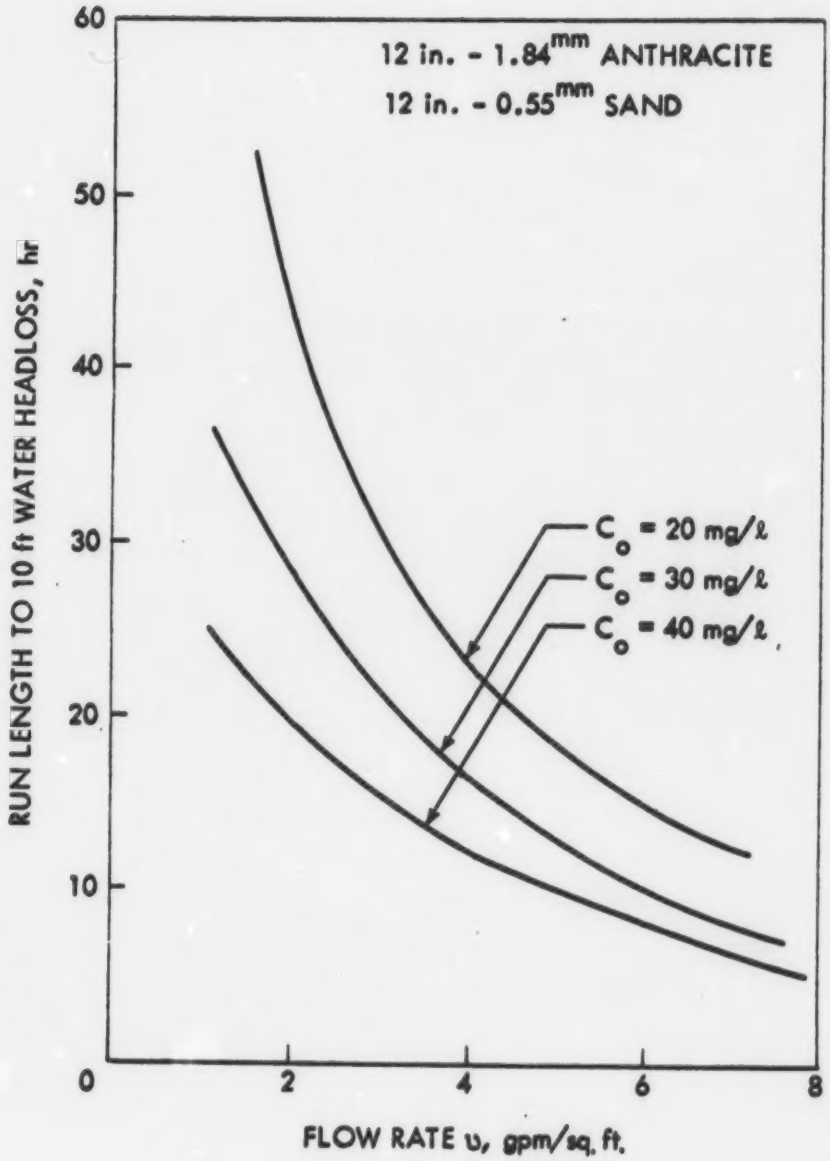


Fig. 15.. Run length vs. flow rate at various influent SS concentrations

hours and the filter might be conservatively designed to operate at 6 gpm/sq.ft. to produce a net filtrate production of 7,440 gpd/sq.ft. at the peak 5-hour flow rate to be expected over the life of the filter.

Higher filtration rates would produce shorter runs and would be subject to operational problems if SS loads or their filtrability varied significantly. The design flow should be based on the average peak flow to be expected during the normal run length.

Without specific data of the type shown in Figs. 14 and 15, this author would not use filtration rates exceeding 6 gpm/sq.ft. based on the peak hourly flow to the filters.

APPLICABILITY OF GRANULAR FILTERS IN PHOSPHATE PRECIPITATION SYSTEMS

General

In wastewater treatment systems involving phosphate precipitation multimedia granular filters would be used to reduce the suspended solids in the waste to a level (less than 2-3 mg/l of SS or level of less than 5 JTU) which would be amenable to subsequent treatment or to final discharge to a waterway. In such applications, the filters could be expected to operate effectively. The chief design problems would involve the following:

- The variable nature of the hydraulic rate of flow through the plant. What is the significant flow rate which should be used for design? Declining-rate filters -- at least without adequate operating data required to establish an operating model for determining optimum operational design -- should be designed for operating at a flow rate not to exceed about 5 to 6 gpm/sq.ft. during the significant total flow rate period.
- The distribution of the total plant flow equally to all banks of filters which operate as a variable declining-rate unit.
- The choice of media sizes and depths and terminal pressure drops to be provided in all filters.

In the proposed flow scheme, declining-rate filters using automatic backwash with the following characteristics should provide the desired filtered water quality with run lengths of about 10-20 hours during the significant flow period:

Q = 6 gpm/sq.ft.

Media = 15 inches of 1.8 mm Anthrafilt
15" inches of 0.55 mm silica sand
or
15 inches of 1.84 mm Anthrafilt
15 inches of 0.77 mm silica sand
6 inches of 0.36 mm garnet sand.

Terminal Head = 10 ft. of water

Hydraulic Design Flow

Although it is important to know the average wastewater flow rate, consideration must also be given to the fluctuations in flow which occur over monthly, weekly, daily and hourly cycles and from dry to wet years. Table 5 shows typical monthly average flow rate variations expressed as a percentage of mean annual flow rate.

Table 5. Variation in monthly wastewater flow rates expressed as as percentages of the average yearly flow. (Ames, Iowa)

Month	Total
January	91
February	90
March	105
April	112
May	121
June	111
July	102
August	99
September	87
October	104
November	91
December	87

Typical variations in daily flow rate are shown in Table 6, Table 7 shows how the hourly flow varies within any one day with maximum hourly flow occurring aroung 10 AM to 2 PM (Fig. 16), probably the significant hydraulic period.

In general, patterns of variation in flow rate are relatively consistent as long as there are no major changes in population characteristics. The above flow patterns (dry weather flows) do not include the excessive flows due to combined sewers.

The maximum hourly wastewater flow reaching a water pollution control plant would be expected to occur on the maximum day of the month of highest flow during a wet year. This maximum flow would be represented by the

Table 6. Variation in daily flow rate expressed as percentages of the average weekly flow rate.

Week day	Total
Sunday	82
Monday	106
Tuesday	104
Wednesday	103
Thursday	103
Friday	104
Saturday	98

Table 7. Variation in hourly flow rate expressed as percentages of the average daily flow.

Hour of day	Percentage	
8 AM	52	
10 AM	142	4-hr pk. = 144
11 AM	150	
12 Noon	145	
2 PM	140	
4 PM	125	
6 PM	115	
8 PM	120	
10 PM	108	
12 Midnight	103	
2 AM	80	
4 AM	50	
6 AM	42	

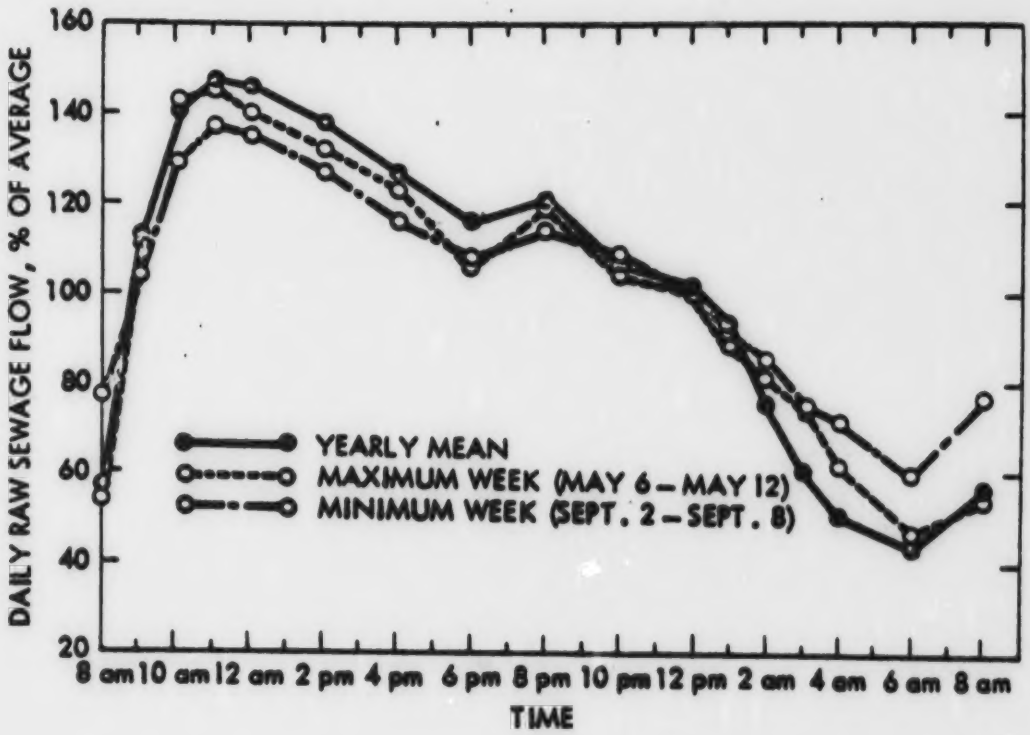

DAILY RAW SEWAGE FLOW, % OF AVERAGE
160
140
120
100
80
60
40
20
YEARLY MEAN
MAXIMUM WEEK (MAY 6 – MAY 12)
MINIMUM WEEK (SEPT. 2 – SEPT. 8)
8 am 10 am 12 am 2 pm 4 pm 6 pm 8 pm 10 pm 12 pm 2 am 4 am 6 am 8 am
TIME

Figure 16. Ranges in <u>typical</u> 24-hr wastewater flow pattern.

product of the average yearly flow rate and the extreme variation in yearly, monthly, daily, and hourly flow rate expressed as a decimal fraction. Then

$$Q_p = \overline{Q} \times \frac{M_p}{\overline{Y}} \times \frac{D_p}{\overline{W}} \times \frac{H_p}{\overline{D}} \times \frac{\overline{Y}_w}{\overline{N}} \qquad (1)$$

where:

Q_p = design peak flow, mgd

$\overline{Q}$ = average yearly flow rate, mgd

$\frac{M_p}{\overline{Y}}$ = ratio monthly peak flow rate to average yearly flow rate (Table 5, May, 1.21)

$\frac{D_p}{\overline{W}}$ = ratio daily peak flow rate to average weekly flow rate (Table 6, Monday Fig. 1.06)

$\frac{H_p}{\overline{D}}$ = ratio hourly peak flow rate to average daily flow rate (Table 7, 4-hr. peak, 1.44)

$\frac{Y_w}{\overline{N}}$ = ratio average yearly flow rate in wet years to average yearly rate for normal years. (1.11 in example city)

Normally, we should not design filters (or any other part of the plant) for the single largest hourly flow rate possible since this flow rate may occur for only few hours in any one year. In normal operation of a treatment plant, the peak flow covers several hours and can be smoothed out by using the incoming sewer line and pumping station storage volume for flow equalization. Consequently, the peak flow rate for a 4-hr. period should be satisfactory for peak hydraulic design purposes, provided filter run lengths at that time are at least 4 hours long.

The relationship between the average and peak 4-hr. wastewater flow

rate for the city represented by the data included herein during wet years would then be

$$Q_p = \overline{Q} \times 1.21 \times 1.06 \times 1.44 \times 1.11$$

$$Q_p = \overline{Q} \times 2.05.$$

The actual peak flow rate which should be used for design purposes must be selected on the basis of the ultimate treatment needs. If short periods of hydraulic overload and reduced treatment efficiency can be accepted, the design flow can be reduced, thereby reducing the size of the treatment units. If such periods of overload cannot be accepted the plant might have to be designed for the most extreme load expected within the design life.

Critical Filter Sizes

In order to be able to distribute raw water and collect backwash water effectively, the maximum size of any one filter should be limited to about 850 sq.ft. This would limit the water filtered by any one filter to a maximum (24-hr. runs) of:

6 gpm/sq.ft. x 850 sq.ft. x 1,440 min/day = 7.35 MGD

In normal practice, at least 4 filters are usually included in a single declining rate unit and thus a single unit would have a maximum capacity of 29.4 MGD (4 x 7.35 MGD). In normal plant operation with a battery of 4 filters, it is unusual when all four filters are in operation simultaneously. In fact, with only 1 hour runs, we find that:

- All four filters can never be in operation simultaneously,
- At most, three filters will be in operation simultaneously only two-thirds of the time,
- In general, only two filter cells are in operation for one-third of the time.

In order to have at least three filters in operation all of the time, run lengths must exceed 1 1/2 hours.

Figure 17 shows the percent of time that all four filter cells will be in operation as a function of run length. This figure says, in effect, that when run lengths are 9.25 hours, the filtration rate on the filters will be 1.33 x the design filtration rate for 20 percent of the time. If run lengths are shorter, the percentage of time of rate overload increases significantly.

With these facts in mind, we can now approach the determination of filter area requirements. We can conclude, therefore, that we have three practical approaches when using 1 variable declining rate unit consisting of 4 filters:

- If normal filter run lengths are expected to be greater than 2 and less than 12-15 hours, filter area would be

$$\frac{Q_{p,\ mgd}}{\text{Design filtration rate, gpm/sq.ft}} \times 1.33 = \text{required filter area.}$$

  This will keep the actual filtration rate at the design rate at all times.

- If normal filter run lengths are expected to exceed 12-15 hours, then the percentage of time all filters are in service approaches 100. Then, the provision of increased filter area is not so important and a lesser percentage increase in filter area is indicated. With run lengths over about 24 hours, there should need be no increase in filter area required.

- If we have actual filter operating data, we can base area requirements on the actual net water production dependable from each sq.ft. of the filter (Table 4).

The critical filter design criteria would then be:

- Use variable declining rate filters,

- Use a minimum of one unit consisting of 4 separate filter cells. (6 on larger units)

- Use a maximum filter size of about 850 sq.ft.

- With larger installations, we would prefer a minimum of 4 units (16 filters) rather than a single unit of large size to provide service in the event one unit (or several filters) needed to be out-of-service for repairs.

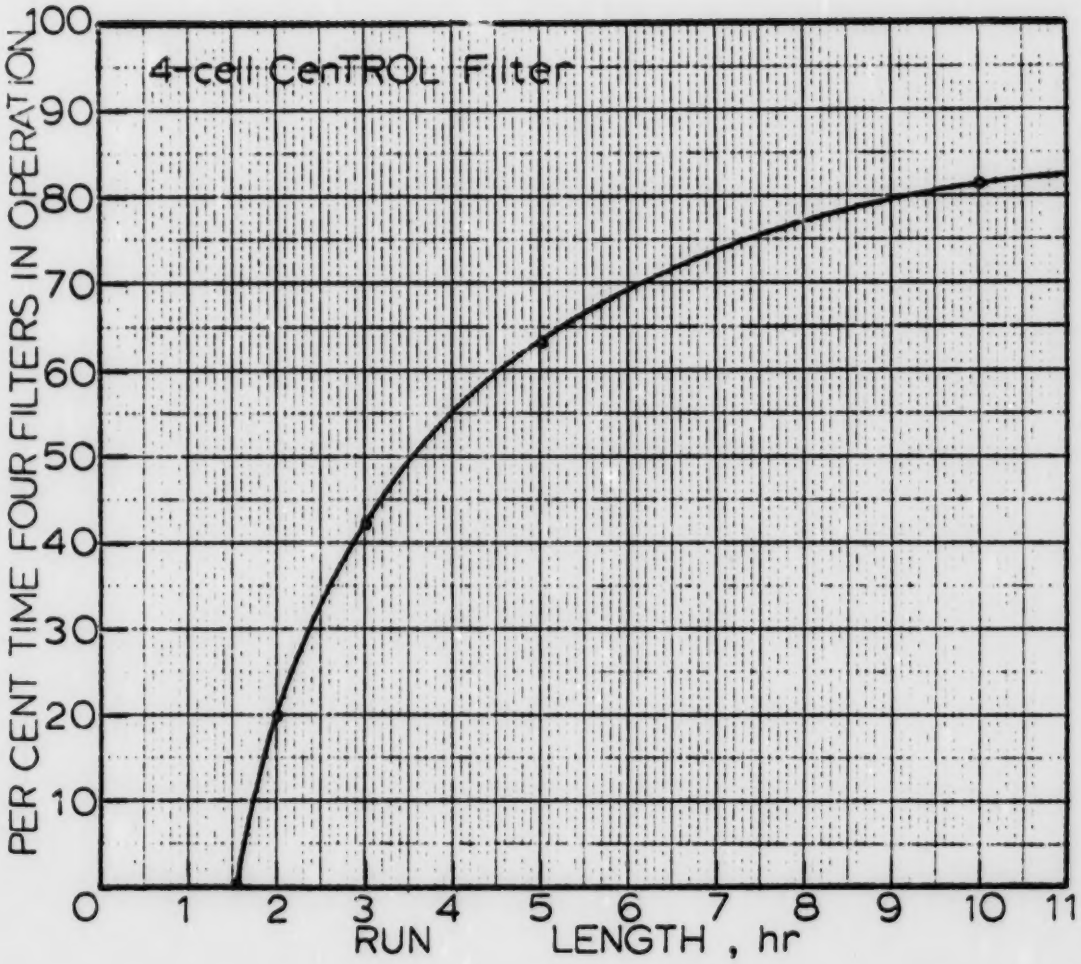


Fig. 17. Per cent of time four filter cells in operation vs. run length

Filter Layout and Details

For maximum construction economy, the 4-filter (or 6-filter) declining-rate units should be rectangular in plan so as to promote common wall construction. In the filtration plant layout, factors to be considered in design include:

- Influent distribution uniformly to all filter units is a critical item in plant design. Where only 3 to 6 units are involved, an overflow-weir type splitter box might be appropriate. Where more units are involved, an influent flow channel might be used along one side of the units with parshall-flume control to each unit. Equal flow distribution is essential.

- An adequate supply of backwash water (to 30 gpm/sq.ft) is essential for proper backwashing. For plants using 3-6 units, no separate backwash pumps need be used if construction typical of Fig. 7 is used. However, all filter units must be connected to the same overflow weir so filters in all units can be used to supply backwash water. (Three filters can only supply 18 gpm/sq.ft. of backwash water to the unit undergoing backwash). To obtain the 20-30 gpm/sq.ft. needed for coarse multi-media filters, additional water is needed from other filtering units. With more than 3-6 units, the filter walls can be reduced in height to reduce total plant cost by using separate backwash pumps and a filter section as shown in Fig. 6a.

- The approximate height of the filtering cell walls would be approximately as follows:

	Without Backwash Pumps	With Backwash Pumps	
Head Loss to Outlet Weir Level	10 ft.	10 ft.	
Weir Level to Wash Trough	4 ft.	0 ft.	
Wash Trough to Media	4 ft.	1 ft.	(Weir level to media)
Media	4 ft.	4 ft.	
Underdrain	1 ft.	1 ft.	
Total	23 ft.	17 ft.	

- The filter design should provide for effective air-water backwash in sequence as follows:

  - Air wash of 4-5CFM/sq.ft. for 3-4 minutes.

  - Water wash of 20-30 gpm/sq.ft. for 10 minutes. (based on media expansion)

- The minimum total backwash cycle would be distributed as follows:

Shut off filter, flow time to drain to trough level	
dumping	5 minutes
filtering	14 minutes
Drain time from trough level to at least 1 foot below trough level	3 minutes
Air wash	3 minutes
Water wash	8 minutes
Return to service	1 minute
Total	17-26 minutes

PROCESS EXPERIENCE TO DATE

Filters are normally designed for the purpose of removing suspended solids, -- and only suspended solids. Thus, filtration applications which are designed for the removal of soluble BOD, for example, are bound to fail. Filters have been and can be applied successfully in tertiary wastewater treatment plants for the following purposes:

- Reduction of SS and BOD from trickling filter and/or activated sludge plants following final sedimentation. Plant efficiency in BOD removal is a function of what proportion of the BOD is represented in the SS.
    - Plants operate w or w/o chemical pretreatment.
- Reduction of SS, their associated BOD, and residual chemical precipitant from trickling filter and/or activated sludge plants or from primary effluents treated with iron or aluminum salts or lime for the purpose of improving solids removal in subsequent sedimentation and for precipitation of phosphates from the plant discharge.
- Since expanded carbon beds can operate effectively with some suspended solids, filters can be used to remove residual biological suspended solids, residual chemical precipitants, and degraded carbon particles from the effluents of expanded carbon beds.

In practice, ample evidence is available to demonstrate that wastewater effluents can be clarified effectively and dependably. It is not difficult to obtain a filtrate that will have a turbidity of less than 5 turbidity units, the maximum turbidity permitted in drinking water according to the U.S. Public Health Service drinking water standards. The difficulty with tertiary filtration plants lies in being able to obtain reasonable, economic filter run lengths which will permit operation with a reasonable consistancy under conditions of both flow variation and waste strength variation. Filters cannot be designed for average flow conditions and be expected to perform adequately under critical flow and solids loading conditions.

Table 8. Tertiary filtration plant locations, Illinois

	Plant Location	Filters	Biological Plant	Remarks
1.	Addison South*	S and L, Graver 3, 10 or 12-ft. Monovalve 5GPM/sq. ft., Air wash-4 CFM/sq. ft., 5 minutes, 24-in. sand, 24-in. anthrafilt, 4800 gals/backwash, 6-8 hour runs.	S and L Oxigest Trickling Filter	Underdrains plugged 2.1MGD Cost - $40,000
2.	Addison North*	S and L, Graver 3, 8-ft. Monovalve	S and L Oxigest	Opened July 1970. Filter design for 3x Mean Flow.
3.	Bensenville	Neptune Micro-Floc	-----	------
4.	Park Forest South	Walker Process 4-gravity filters w/air wash	-----	------
5.	Lisle (1971 completion)	Eimco 6, 12-ft. diameter, dual valve, air wash	Extended aeration	Multi-media
6.	Lake Zurick	Eimco 4, 8-ft. filters, air-wash, 5GPM/sq. ft.	Walker Contact-Stabilization plant	Very short runs. Additional filters to be added to reduc rate.
7.	Barrington	Neptune, Micro-Floc	-----	------
8.	Hanover Park (M.S.D.) (experimental)	Walker Process	Follow Dorr Clarifier-flocculators	Experience Published (4)
9.	Woodale South	S and L, Graver	-----	------
10.	Glendale Heights (contract awarded)	S and L, Graver	-----	------
11.	Lake County Public Works Dept. (not built)	Eimco	-----	------

* Supt. Anthony LaRocca

Supposedly, we should be able to learn from the experience of others. To summarize the experience of others, Appendix A was prepared as a summary of the published experience with Granular Filters for Wastewater Filtration. This is not a complete list, since there should be a significant number of EPA demonstration project reports with more complete operating data for more recent American experience. It is safe to conclude, however, that European experience with and use of filtration for improving wastewater quality is more extensive than American experience.

In view of the extensive published experiences, we might expect that current tertiary filtration experience would be emminently successful. Unfortunately, it is not. To the authors personal knowledge, there are no tertiary filtration plants yet built in Iowa and none in any significant numbers in any surrounding states except in Illinois. Illinois plants have been designed in accordance with the (Illinois) Sanitary Water Boards Technical Release 20-24 which requires that wastewater discharged to a stream with less than 1:1 dilution have a BOD less than 4 and suspended solids less than 5 mg/l. To approach these levels, Illinois requires filtration of the effluent prior to discharge as the first stage treatment in a program of improving stream water quality. As of about May, 1970, for example, the 5-county area (Cook-DuPage-Kankakee-Lake, and Will Counties), near Chicago had approved 11 such tertiary filtration plants, about 1/2 were in operation, none of them really successfully. The meager data on these is summarized in Table 8.

In early 1970, Anthony LaRocca, Supt. of the Addison plants gave an oral report at an operators meeting indicating the type of operation secured at the Addison South plant. SS applied to the filters averaged 45-50 mg/l and varied from 35-90 mg/l. With applied SS of 35 mg/l, the

effluent carried from 2-10 mg/l. With applied SS of 90 mg/l, the effluent contained from 14-27 mg/l. With applied BOD of 44 mg/l, the effluent was at 20 mg/l.

Other tertiary treatment plants know to be in operation include:

- General Filter Company (Ames, Iowa) plant at:

  Tree Top (General) Apartments
  Atlanta, Georgia

  This is a 98 gpm dual valve-dual media fully automatic plant with air wash operating on an apartment complex waste treated using an extended aeration package plant. Plant performance reported to be excellent.

- Walker Process Plant at Atlanta, Georgia

  Dual-media gravity filters for filtration of extended aeration effluent. Unsatisfactory performance, since heavy loads bypass plant into chlorine-contact tank from which filter backwash is drawn. Media-retaining strainers clog with backwash SS.

- General Filter Company

  C.D. Hulten Engerprises
  Woodbridge, Virginia

  A tertiary filtration plant at Dale City, Virginia following lime precipitation of phosphates following biological secondary treatment. Uses GFC Contra-Flo units and CenTrol filters for treatment rate of 2100 gpm (3 MGD). Plant privately owned. Plant to start in operation almost immediately.

- Eimco Corporation
  St. Lake City, Utah

  Under WQO, EPA Contract WP-14-12585, Eimco is operating a 100 gpm pilot plant for physical-chemical treatment including use of

powdered activated carbon. Tertiary filters (0.9 mm coal, 0.5 mm sand at 2-3 gpm/ft^2) using dual media are used. Backwash with air at 4-5 CFM/sq.ft. for 1 minute, air-water backwash for 1/2 minute, and water backwash at 25 gpm/sq.ft. for 5 to 8 minutes. Results reported to be excellent.

- District of Columbia Water Pollution Control Plant
  Washington, D.C. (30)

The WQO-EPA is operating an advanced waste treatment pilot plant consisting of single-or two-stage lime precipitation followed by filtration. Lime precipitation usually removes 90 percent of the phosphorous and 55 percent of the organic carbon. With efficient clarification using ferric flocculant, dual media filtration at 3 gpm/sq.ft. removes an additional 5 percent of the phosphate in a 2-stage lime process and produces a residual turbidity of about 0.5 J.T.U. (Table 9). Filter runs averaged 59 hours when reducing turbidity from 4.9 J.T.U. to 0.5 J.T.U. Backwashing required a 20 gpm/sq.ft rate (10 minutes). The filter media included 18 inches of Anthrafilt (size unspecified) and 6 inches of 0.40 to 0.45 mm sand. A sand filter containing 20 inches of 0.40 to 0.45 mm sand produced effluent quality equivalent to the dual media filter, but run lengths averaged only 32 hours. Except for 2-stage operation with ferric addition, the filters did not remove the residual colloidal "haze" resulting from phosphate precipitation with lime and effluent turbidities were in the range of 2.4-3.3 J.T.U. Finer sand media (0.30 effective size) reportedly did not remove the "haze" and produced excessive head losses.

A later paper (31) indicated that the dual media filters con-

Table 9

Plant Performance
with
Two-Stage Lime Precipitation (3)

Process	Total Phosphorus		TOC		Turbidity
	Residual mg/l	Removal (4) %	Residual mg/l	Removal (4) %	Units
No Additives (1)					
Clarification	1.00	87	19.5	45	7.7
Filtration	0.78	90	16.5	53	3.3
Carbon Adsorption			6.5	84	3.5
Ferric Chloride (2)					
Clarification	0.55	93	14.1	64	4.9
Filtration	0.39	95	10.1	74	0.5
Carbon Adsorption			2.3	95	0.4
Polymer A23 (3)					
Clarification	0.54	91	15.7	55	5.3
Filtration	0.39	94	12.9	65	2.4
Carbon Adsorption			3.2	92	1.6

(1) Rise rate 1500 gpd/ft^2; temperature 14°C; pH 11.8; length of run 50 days; lime 340 mg/l CaO

(2) Rise rate 900 gpd/ft^2; temperature 20°C; pH 11.5; length of run 56 days; lime 290 mg/l CaO; Ferric 6-12 mg/l Fe^{+++}

(3) Rise rate 1060 gpd/ft^2; temperature 23°C; pH 11.4; length of run 16 days; lime 260 mg/l CaO; Polymer 0.15 mg/l A-23 (anionic polymer by Dow Chemical Company)

(4) Accumulated removals from secondary effluent

tained 18 inches of 0.9 mm anthracite (no u.c. given) on top of 6 inches of 0.45 mm sand (no u.c. given). P-C treatment reduced the raw waste phosphorous level from 8.4 mg/1 to 0.27 mg/1 after lime precipitation and to 0.18 mg/1 after filtration. Conventional tertiary treatment reduced the raw waste phosphorous level to 6.7 mg/1 in secondary treatment, to 0.34 mg/1 by lime precipitation, and to 0.14 mg/1 by filtration.

Another pilot plant at the same location (32), treated primary effluent by 4-stage step aeration with alum addition in the final aeration pass for phosphorous removal followed by secondary clarification and multi media filtration for tertiary solids removal. Raw phosphorous levels were about 8.15 mg/1. Dual media filters (18 inches of 0.90 mm coal on 6 inches of equal quantities of 0.40-0.45 mm and 0.8-1.20 mm sand) and a tri-media filter (25 inches coal on top of 10 inches of sand and 5 inches of garnet, Microfloc designation MF197) were loaded at 2.4 gpm/sq.ft. and regularly produced filter runs of 24 to 32 hours. The tri-media filters had operating results as follows:

Month	Waste	SS, mg/1	P, mg/1
October	Secondary Effluent	41	2.2
	Filter Effluent	9	0.58
November	Secondary Effluent	50	2.9
	Filter Effluent	16	0.97
December	Secondary Effluent	33	1.9
	Filter Effluent	8	0.65
January	Secondary Effluent	26	1.6
	Filter Effluent	7	0.61

The dual media effluent quality was not as good. No filter operating characteristics (run lengths, etc.) were described.

CAPITAL AND OPERATING COSTS

The costs involved in tertiary filtration of wastewaters are even more difficult to obtain than operating experience data. Usually, filtration cost data are buried in total general contract bidding and are not too available. The main item of variation in the cost of a variable, declining rate filter of the CenTROL type is in the cost of the construction of the filter plant and assembly of the equipment. The contractor usually bids a general contract using different unit prices for concrete in-place depending on plant location. A unit price of $200/cu.yd. for concrete in place was assumed in order to estimate filter construction costs. Costs included were:

Filter floor

Underdrain slab

Effluent box

Filter wall and cell walls

All equipment

Profit (10%)

Legal and engineering fees (20%)

Three plant flow capacities (1, 5, 10 MGD) and three filtration rates (2, 4, and 6 gpm/sq.ft) were used. In each case, the actual filter area used was increased 33.3 percent to compensate for filter down time for backwash. The relationship between plant flow capacity at each filtration rate and capital cost is shown in Fig. 18. A published capital cost comparison is shown using data from Weber et. al. (33).

Fig. 19 shows the sensitivity of the capital cost to change in filtration rate for a plant capacity of 10 MGD. The figure indicates that

substantial capital costs can be saved by use of higher filtration rates. Similar cost data indicate that very little unit cost savings is effected as plant size increases above about 60-80 MGD, since the larger size plants merely require a much larger number of the maximum size filter unit.

Regretfully, operating cost data are also lacking. Lynam and Bacon (34, 35) estimate the total operating cost of a 10 MGD performance capacity plant based on Hanover plant data to be 2.40 cents/1000 gallons. Cost data given were:

Capital cost	$ 240,100	(Fig.18)
Amortized cost/MG	3.804	
Operation and Maintenance/MG	13.009	
Chemical/MG	6.990	
Power	0,202	
Total	$ 24.00/MG	

This included prechlorination costs. Data available to the author would appear to indicate that an average operating cost for tertiary wastewater filtration should be in the neighborhood of 2.0-2.5 cents/1000 gal. Naylor (4) reported that the cost of conventional downward-flow sand filters operated for 15 years at Luton, including depreciation and running costs, including repayment of capital cost, worked out at 5.45 cents/1000 gallons. They expected upward flow filters to significantly decrease this operating cost.

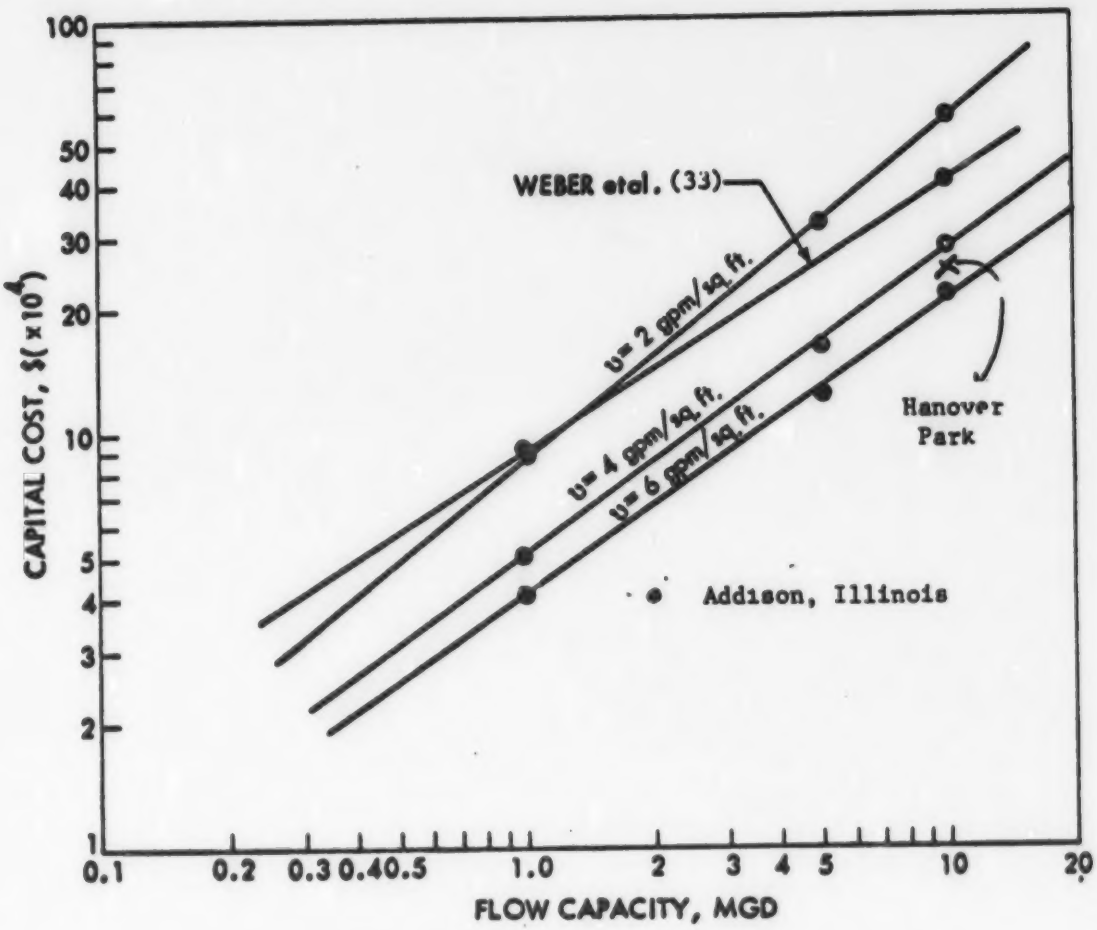


Fig. 18. Capital cost vs. designed flow capacity for various flow rates

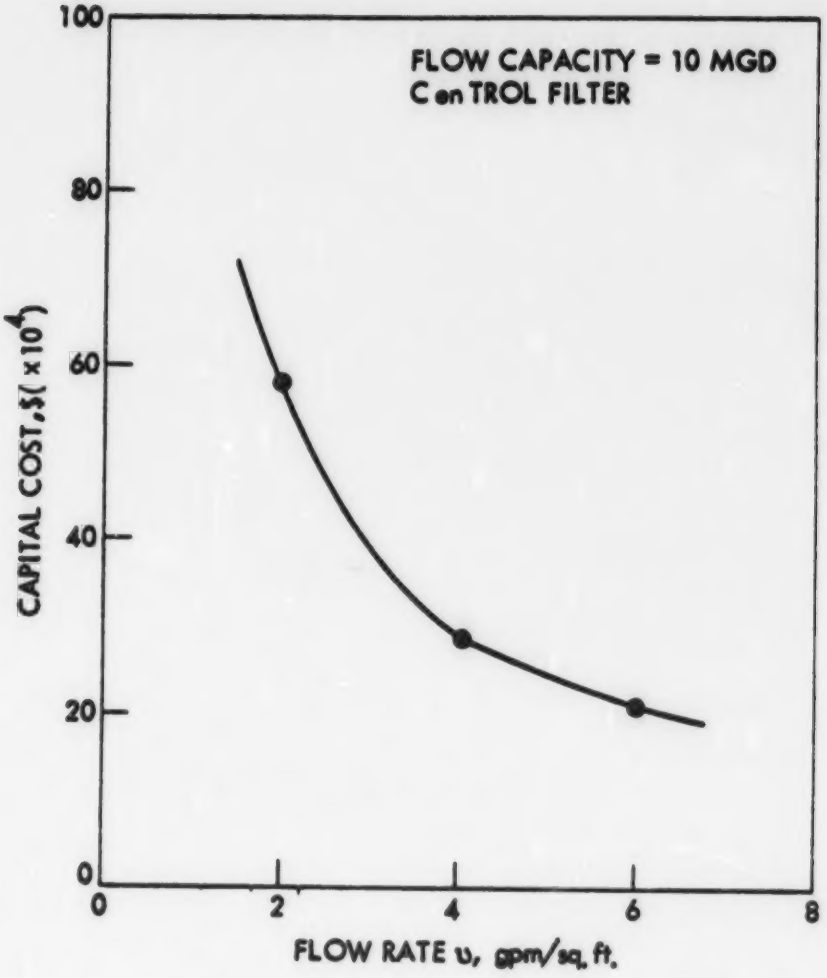


Fig. 19. Capital cost vs. filtration rate for a designed flow capacity of 10 MGD

SAMPLE CALCULATIONS AND TYPICAL PLOTS

Filter design characteristics

In the design of a filter for tertiary waste treatment, the filter characteristics which must be established include:

Filter media size and depth
(single media; graded or ungraded)
(multi-media; types, specific gravities, sizes)

Filtration rate, gpm/sq. ft.

Terminal pressure drop, ft. of water

Present design practice tends to use filter media which is too small in size and too shallow in depth, filtration rates and terminal head losses which are too low, and filtration experience which is based on potable water rather than waste filtration experience. As a result field experience available in the United States is not really sufficient to outline with the required degree of precision expected filter performance under the wide variety of waste filtration applications that are possible in tertiary treatment situations. In those situations where the actual waste to be treated exists, a series of pilot plant studies conducted over a period of several weeks can provide the information needed for optimum economical design purposes.

In designing a pilot plant study, the following points should be considered:

1. Filters should be compared on the basis of their ability to provide equivalent performance. Two or more operationally optimum filters may be said to provide equivalent performance when they produce the same quantity and quality of filtered water from the same water source during the same time period. Thus, equivalent performance

means that the quantity and quality of the filtered water produced in one day, for example, must be the same, but their media size, media depths, filtration rates and terminal head losses may be significantly different.

2. The filter which permits the equivalent performance to be achieved at least cost would represent the optimum filter design. In general, few, if any, filters have been designed which can produce equivalent performance at least cost. A recent paper by the author (2), indicates that least cost filtration of an iron suspension (Figure 12) tends to require coarse media, high filtration rates, and relatively large media depths. In Figure 12 the least cost filter design is not reached, but must be for media size greater than 1.3mm and a correspondingly higher filtration rate and media depth. Optimum design in this case provided about 30% cost savings over that obtained by standard practice design.

Ives (1) in a very recent paper, has also suggested techniques for least cost filter design. In applying these techniques using two mathematical models for iron removal and British filter cost data, Ives concluded that the least cost filter would have the following characteristics:

Media depth:	8 ft.
Media size:	1.5mm
Head Loss:	10 ft.
Filtration Rate:	7.5 gpm/ft^2

These data would tend to confirm those by Huang and Baumann (2) Ives (37) has also demonstrated that for each media grain size there is a maximum media thickness beyond which no operational optimum can be

established and provision of media depth beyond this maximum merely wastes filter media and filter space and adds slightly to operational costs.

3. The length of the operational optimum filter run desired will depend on the method of filter operation -- whether manual or automatic backwash and filter operation is provided. More water will be used for backwash as the frequency of backwash increases. With manual operation, filter runs should exceed 24-30 hours. With automatic operation, shorter filter runs are economically feasible, but ordinarily filter run length should exceed 12-15 hours.

With these considerations in mind, a filter pilot plant can be designed and operated to provide the data necessary to determine the optimum operational characteristics and to approach the optimum economical characteristics of a tertiary waste treatment filter. Such a filter pilot plant might involve the operation of three or four batteries of 4 filters per battery (diameter at least 50 x largest filter media size to eliminate scale effect) in a series of runs. The filters in each battery would be filled with 2", 8", 16", and 32" of filter media. The run characteristics might be designed in accordance with the following phases:

Phase A Study of Single-Media Size Effects on Performance.

Filter Batteries A, B, and C contain media sizes 1, 2, and 3 respectively.

Filter runs 1, 2, 3, etc. using all batteries at flow rates of 2, 4, 6, 8 gpm/ft^2 respectively, observing effluent quality and head loss in each filter.

Phase B Using Media Size from Phase A which promises desired operational characteristics, study effect of flow rate on performance.

Filter Batteries A, B, and C all contain same media size indicated in Phase A.

Filter runs 1, 2, and 3 made using batteries A, B, and C operated at filtration rates of 2, 4, 6, or 4, 6, and 8 gpm/ft^2 respectively.

Phase C Study of single media filters vs. dual (or multi-) media filters using media size and flow rates which promises desired operational characteristics observed in phases A and B.

Filter Batteries A, B, and C contain different media.

Battery A - Samed media as Phase B.

Battery B - Top media, coal 2 x media size in Battery A, bottom media same as Battery A.

Battery C - Top media, coal 1-1/2 x media size in Battery A, bottom media same as Battery A.

Figs. 20 and 21 show a schematic diagram of the pilot tertiary wastewater filters and a schematic diagram of an individual filter cell (the filter shown in a cell containing 12 inches of anthracite and 12 inches of sand).

Figs. 22 through 31 show some of the results of a series of pilot plant runs made with such a battery of filters in the removal of suspended solids from a typical trickling filter plant final effluent at Ames, Iowa. Three media sizes were used (passed/retained, 18/20 or 0.92 mm, 25/30 or 0.65 mm, 30/35 or 0.55 mm) in phase A runs, Fig.22 shows the suspended solids in the influent and effluent and head losses observed in a run at 2 gpm/sq.ft. The following conclusions can be drawn from these results:

- The three filters gave approximately the same effluent quality, with the 0.55 mm sand providing the best quality, 0.65 sand the next, and 0.92 mm the next best.

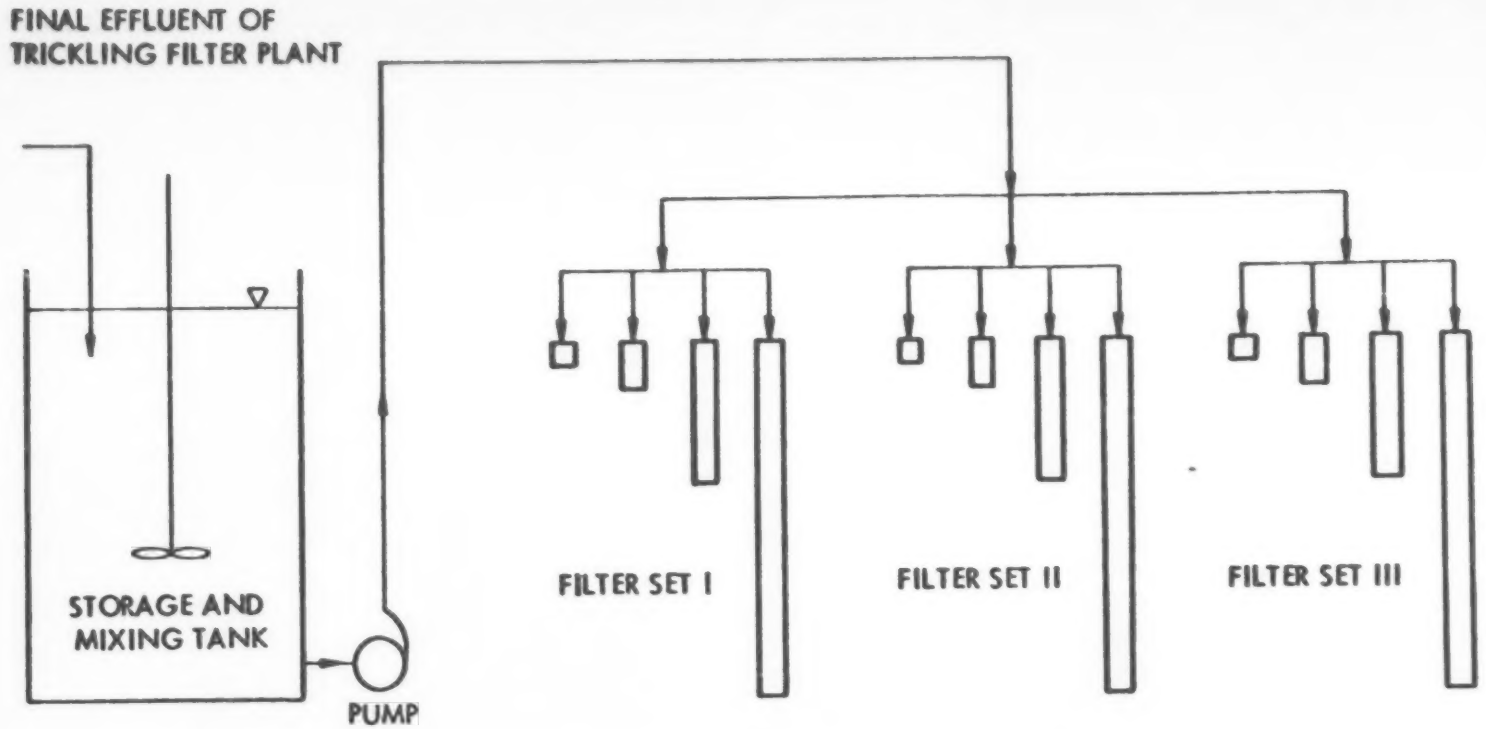


Fig. 20. Schematic arrangement of pilot tertiary wastewater filters

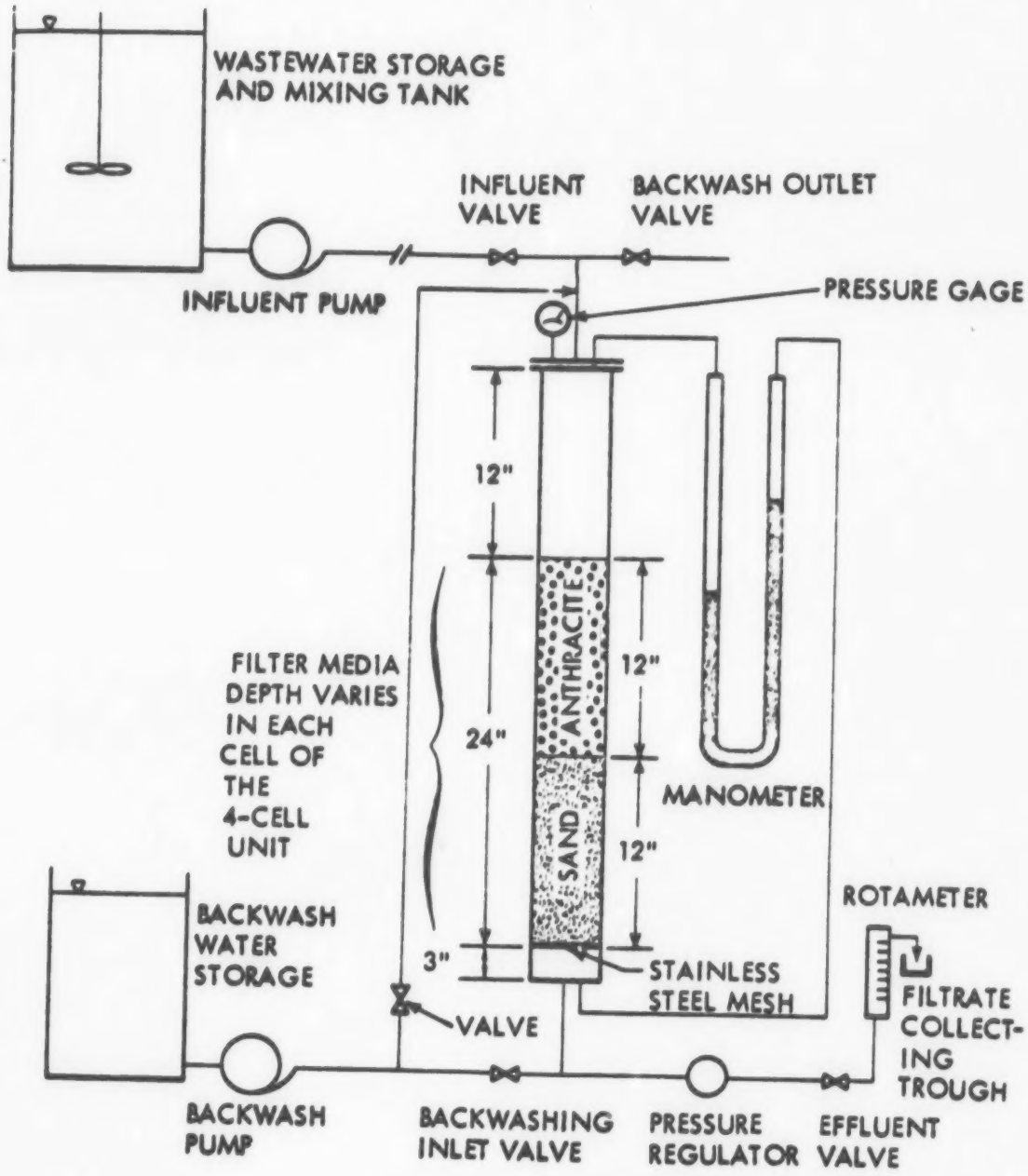


Fig. 21. Schematic diagram of individual filter cell (four cells in each filter set)

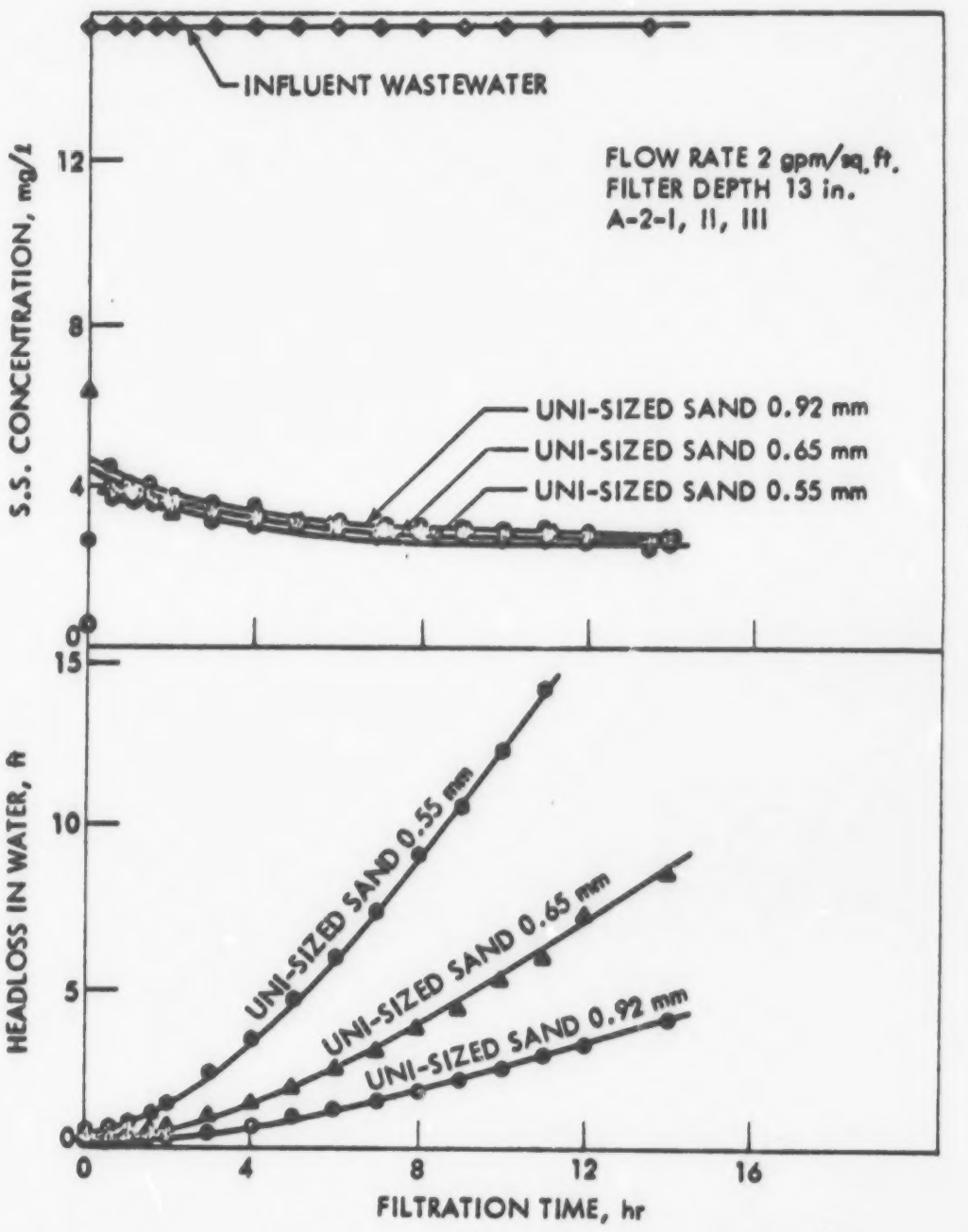


Fig. 22a. Effluent quality and headloss vs. filtration time at various sand sizes

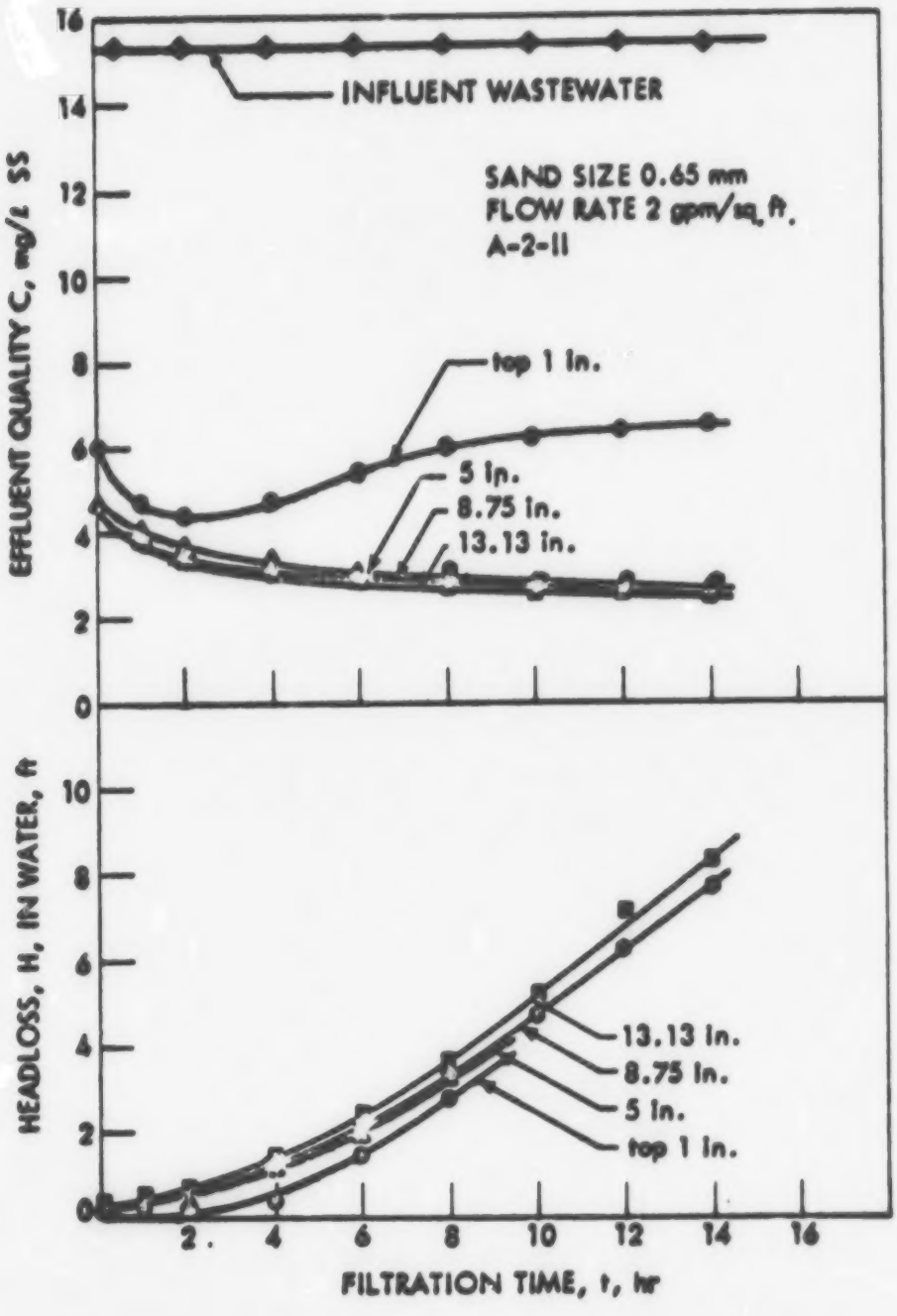


Fig. 22b. Effluent quality and headloss vs. time for a single sand filter

- All three filters provided run lengths of at least 8 hours under a 10 ft. head loss at the filtration rate of 2 gpm/sq.ft. A similar run at 6 gpm/sq.ft. produced only a 2 hr. run at a 10 ft. head loss.

Fig. 22b shows the results from the individual filters in the 0.65 mm sand filter battery at the 2 gpm/sq.ft. filtration rate. These data indicate that most of the head loss and most of the solids removal is in the top inch of the media. This means that surface removal is a probable occurance with this media size. The curves in Figs.22a and 22b are typical of those which can be generated from the operation of the pilot filters.

Fig. 23 shows the various filter operating characteristics in one Phase B run made at filtration rates of 2, 4, and 6 gpm/sq.ft. using 0.92 mm sand. The figure shows the results of one run when the influent suspended solids were at 34 mg/1. The results lead to the following conclusions:

- The effluent quality is best with the lowest filtration rate (5 mg/1) and becomes poorer with increased filtration rate.
- The head loss is least, as might be expected, for the 2 gpm/sq.ft. filtration rate, and increases with the 4 and 6 gpm/sq.ft. rates.
- With this filter media, a filtration rate of 2 gpm/sq.ft. and a media depth of 30-40 inches could be expected to provide excellent suspended solids removal and a 6-hour filter run with a 10 ft. terminal head loss.
- Undoubtedly, this coarse media is not satisfactory for the given application.

Accordingly, several Phase C runs were made using filter media characteristics as follows:

Filter Battery 1, 16/18 sand (1.09 mm)

Filter Battery 2, Top: 12 inches of 10/12 coal (1.84 mm)

Bottom: 12 inches of 20/25 sand (0.75 mm)

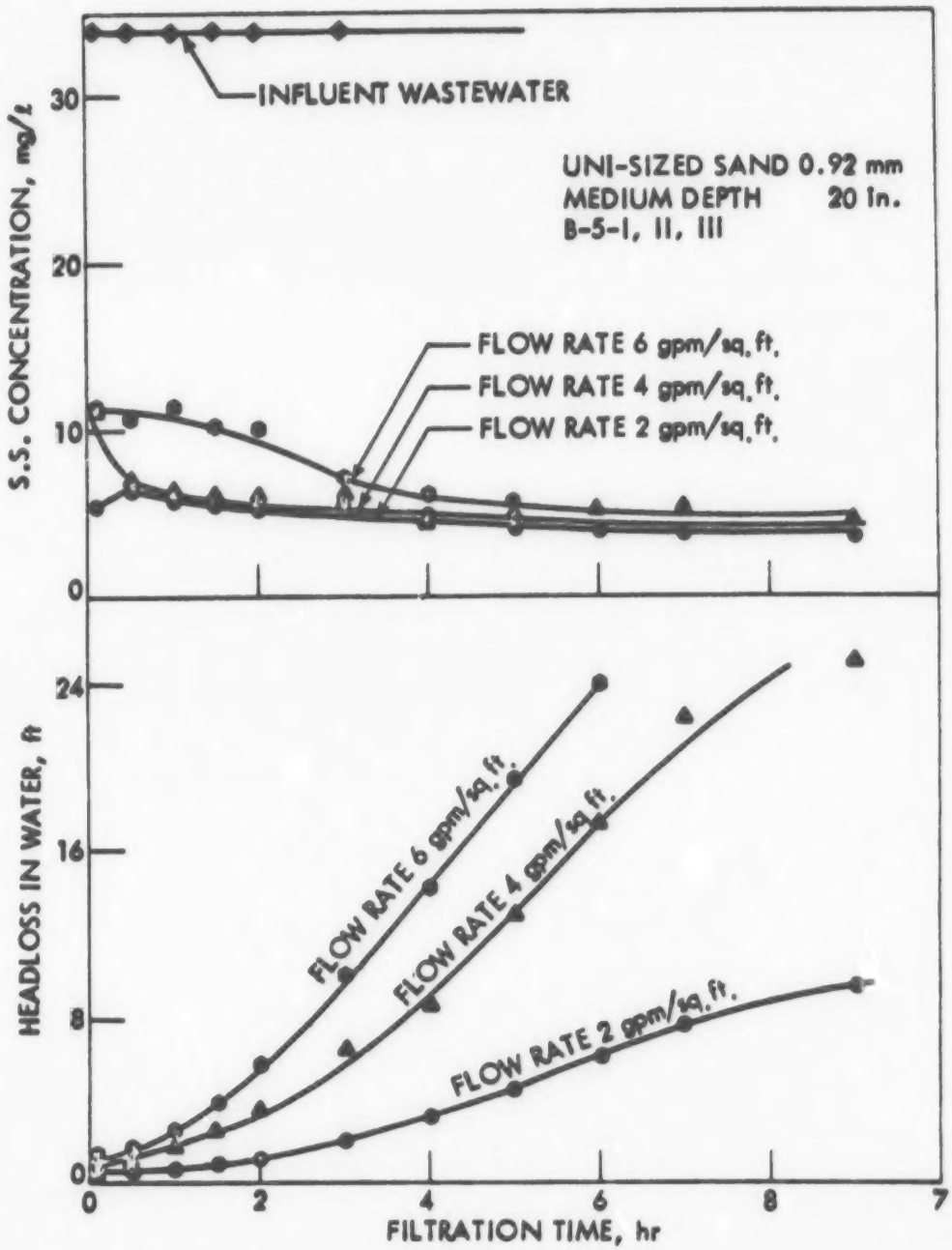


Fig. 25. Effluent quality and headloss vs. filtration time at various flow rates

Filter Battery 3: Top: 12 inches of 16/18 coal (1.09 mm)

Bottom: 12 inches of 20/25 sand (0.75 mm)

Figure 24 shows the results of a run at 4 gpm/sq.ft. when the influent SS were 47 mg/l. The results lead to the following conclusions:

- The three filters provided a different quality of effluent:
  1.09 mm sand -- 7.8 mg/l SS
  1.84 mm coal - 0.77 mm sand -- 6.7 mg/l SS
  1.09 mm coal - 0.77 mm sand -- 5.5 mg/l SS

- The 1.84 mm coal-0.75 mm sand filter provided the lowest head loss, such that a 12-hour run could easily be obtained with a head loss of well under 10 ft. at a filtration rate of 4 gpm/sq.ft.

Figure 25 shows the relationship between suspended solids as related to filter depth after various periods of filtration using the data for the run whose head losses are shown in Fig. 24. These data indicate that:

- The sand filter shows a progressive movement of solids deeper into the filter and demonstrates that the filtrate quality continued to increase with increasing depth.

- With the coal-sand filters, there did not seem to be any major improvement in filtrate quality with added sand depth. The sand-coal intermix zone appears to be the point where maximum effluent quality is attained.

Figure 26 shows the results of a run made to evaluate the effect of using different coal size (1.09 mm and 1.84 mm coal) on top of 0.77 mm sand. In each case, the relationship between head loss and time is represented by an exponential curve. The removal appears not to be an adsorption removal around the media in this case, but appears to be a bridging and/or straining which occupies pores. Thus, the deposit is subject to compression with consequent porosity change and an increasing rate of head loss buildup with time. In general, the use of the 1.84 mm coal-0.77 mm sand filter provided reasonable run lengths at all filtration rates with solids levels in the 30-50 mg/l range.

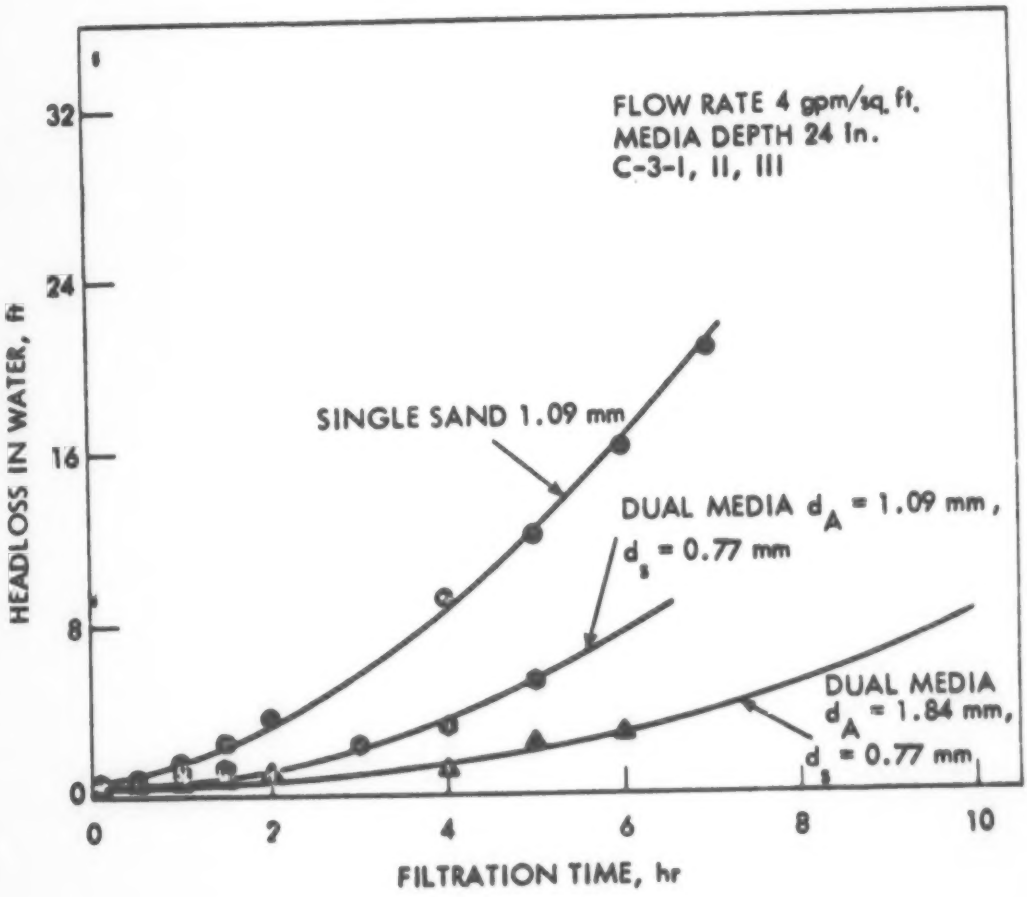


Fig. 24. Comparison of headlosses of single and dual media filters

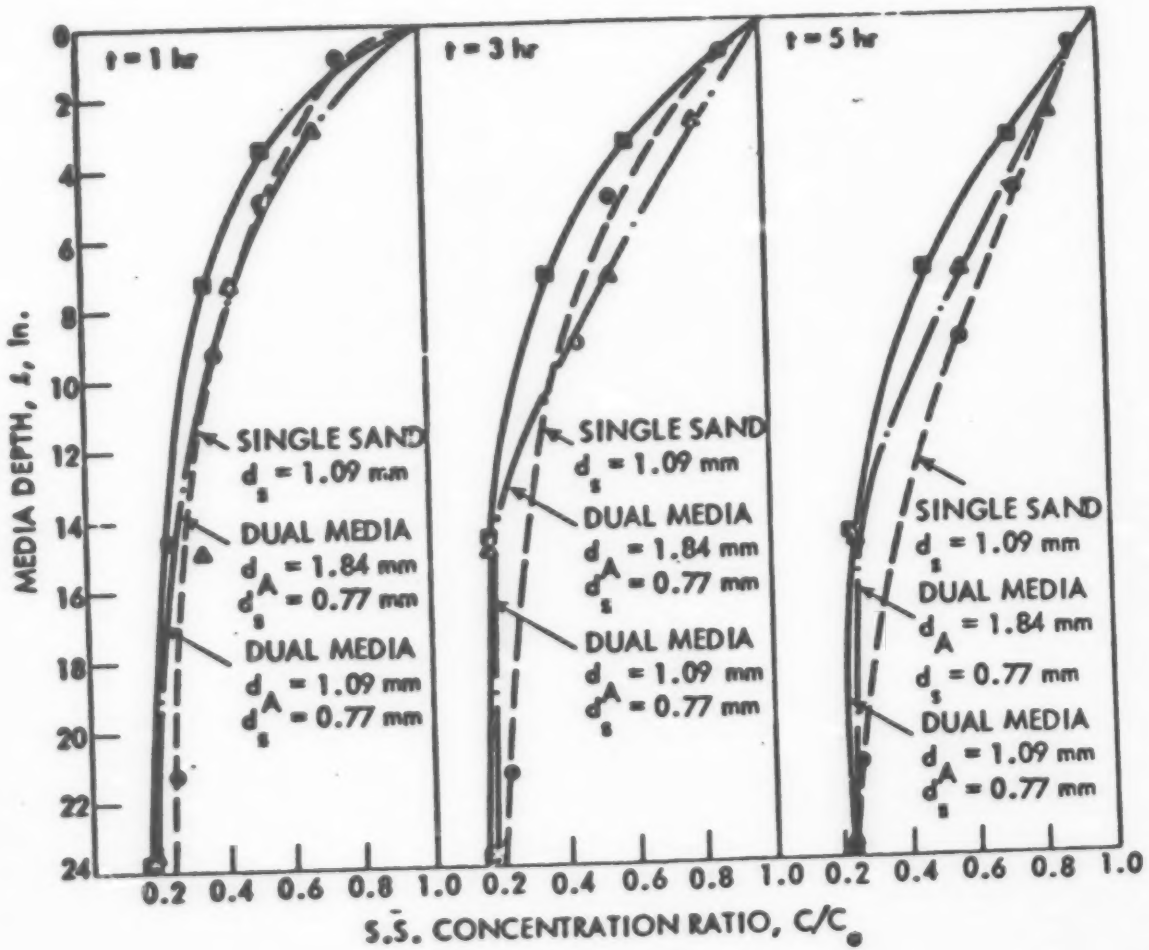


Fig. 25. Comparison of removal efficiencies of single and dual media filters at flow rate of 4 gpm/sq. ft.

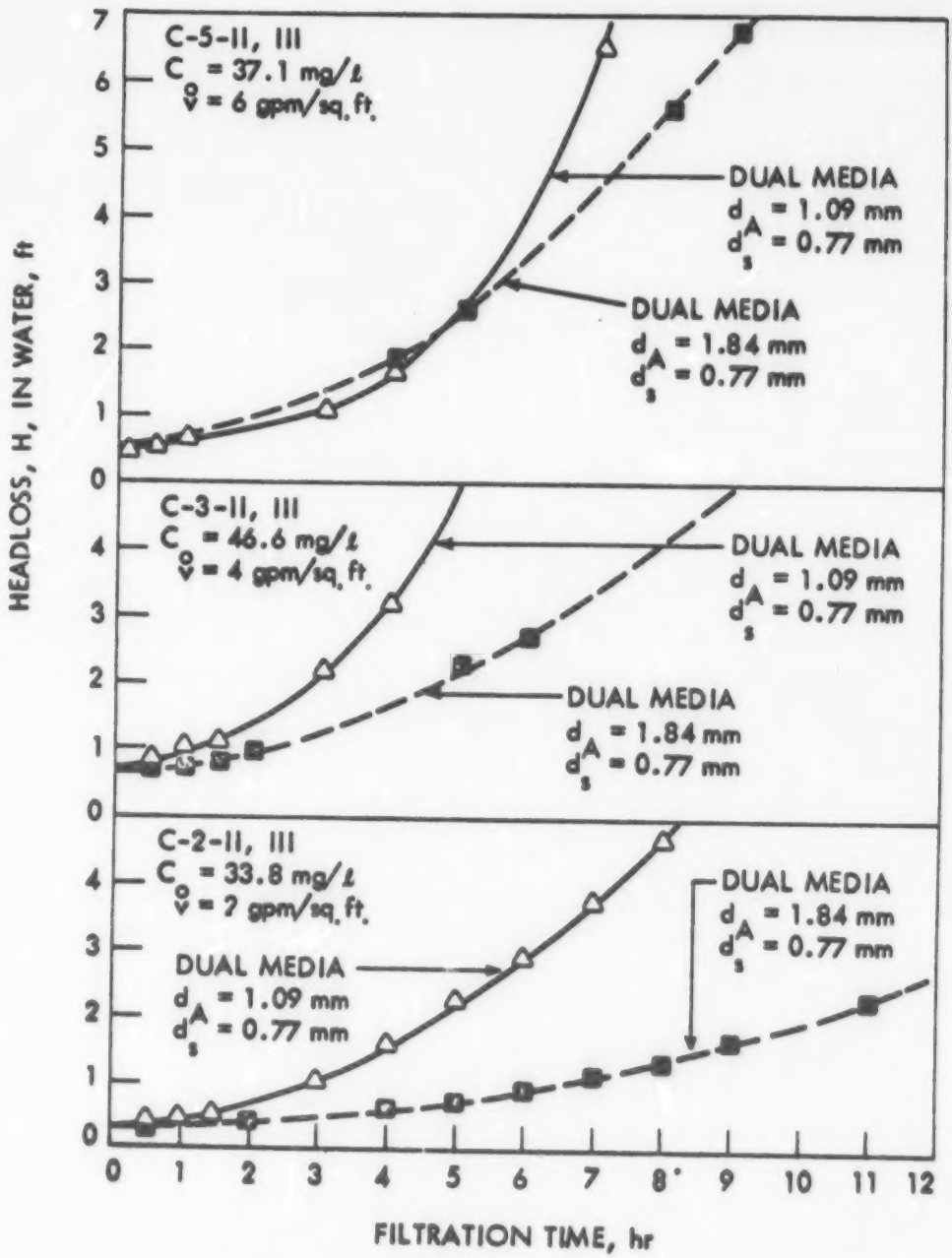


Fig. 26. Comparison of headlosses of various anthracite sizes at different flow rates through the 24 in. full depth filter

Figure 27 shows the results of a series of runs at flow rates of 2, 4, and 6 gpm/sq.ft. using three filters containing 24 inches of 1.09 mm, 1.30 mm and 1.84 mm coal. This series was conducted to determine what depth of coal should be used in a dual media filter. Figure 27 demonstrates that with influent solids of 7.0, 16.1, and 21.6 mg/l there was little additional solids removal in the coal below a coal depth of about 12 inches. In all cases, the 1.84 mm coal provided significantly longer runs. Therefore, this series led to adoption of 12 inches of 1.84 mm coal as the top media in subsequent runs, and a recommendation for the use of 15 inches of coal to provide for coal-sand intermix in any full-scale plant.

The filter battery was then used to determine the effect of sand size and depth used as the lower media in dual media filters. Sand sizes of 0.55, 0.65, and 0.77 mm were used (12 inches) under 12 inghes of 1.84 mm coal. Fig. 28 shows the removal of suspended solids with depth after filtration times of 5, 9, and 53.5 hours of filtration when the filters were operated at 2, 4, and 6 gpm/sq.ft.

The figure indicates coal-sand intermix zones as follows:

1.84 mm coal--0.77 mm sand ----------	2 inches
1.84 mm coal--0.65 mm sand ----------	4 inches
1.84 mm coal--0.55 mm sand ----------	6 inches

Table 10 lists the pertinent data from this run. In general, the finer sand media gave slightly better effluent quality, particularly at the higher filtration rate. In general, effluent turbidity was in a satisfactory range. Run lengths with ten feet of head are also given, indicating only slightly reduced run lengths using the 0.55 mm sand. The results also indicated that little quality improvement occurred in the sand depths below the coal-sand intermix zone. Since the dual media filter using 0.55 mm sand required a lesser backwash rate than the other media (Fig. 10), sub-

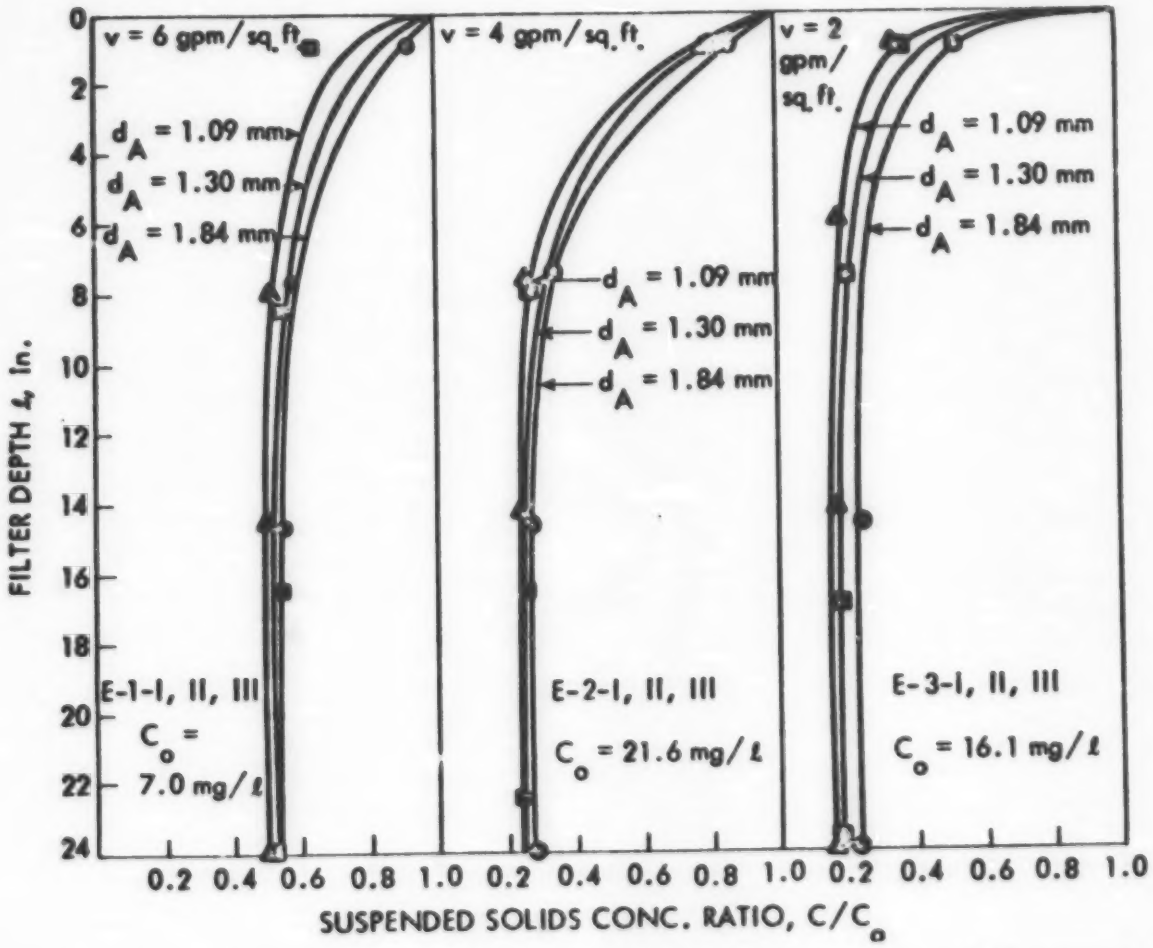


Fig. 17. Effect of single media anthracite size on solid removal efficiency at t = 11 hr.

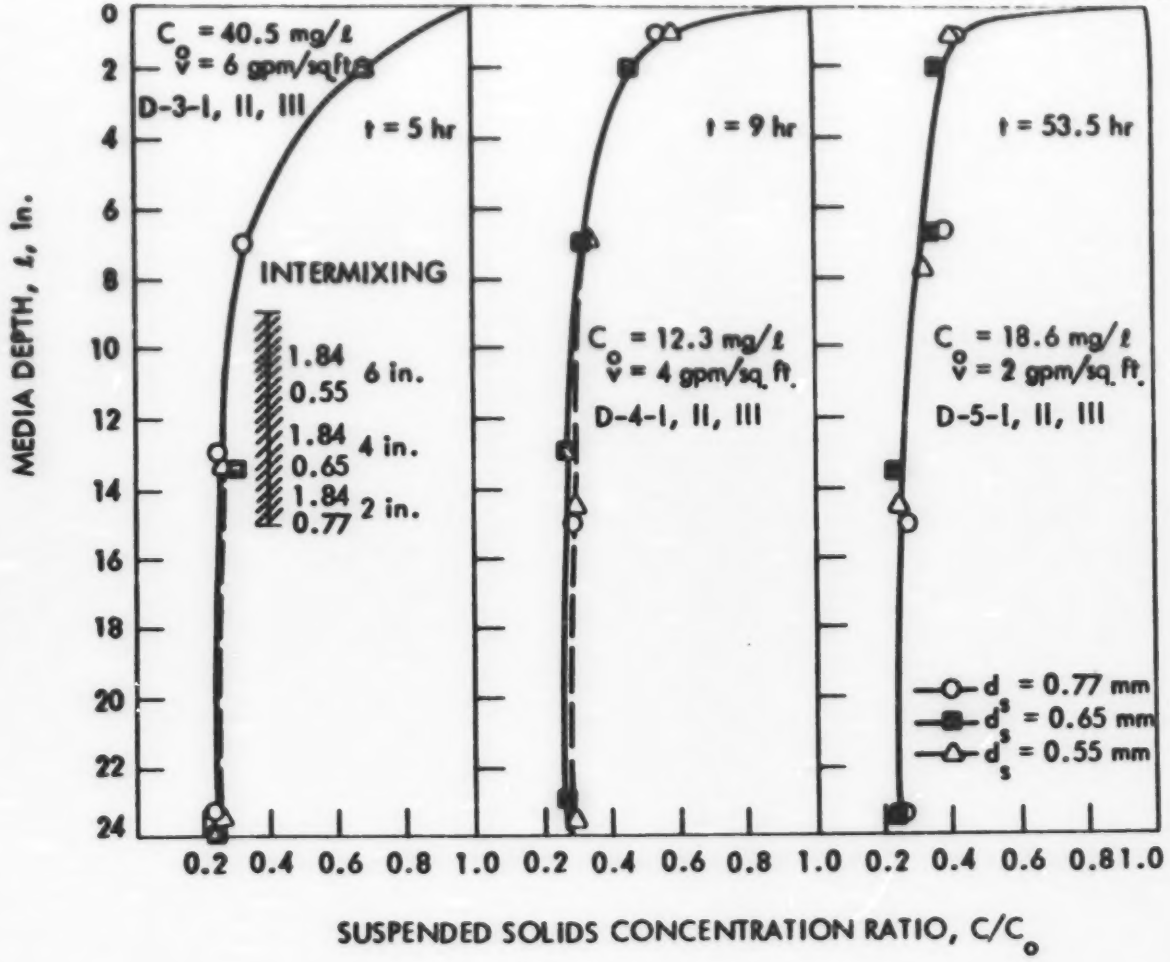


Fig. 28. Comparison of removal efficiencies of various sand sizes at different flow rates

Table 10. Physical characteristics and operating conditions of dual media filters - varying media size of sand (phase D runs)

Run no.	C_o, SS mg/l	Rate gpm/ sq.ft.	Filtrate C*, SS mg/l media sizes, mm 1.84[a] 0.77	1.84[b] 0.65	1.84[c] 0.55	Headloss ft. water	Run length* to given headloss, hr. media sizes, mm 1.84[a] 0.77	1.84[b] 0.65	1.84[c] 0.55
1 **									
2 **									
3	40.5	6	5.4	5.2	5.0	10	9	5.5	7
4	12.1	4	2.8	2.8	2.8	10	28	25	25
5	20.0	2	2.8	2.8	2.6	10	60	46.5	52

* Representative filtrate quality from 4th filter cell (24 in. depth) of each filter set.

** Mixer out of order.

[a] 1 in. intermixing.

[b] 3-4 in. intermixing.

[c] 4-6 in. intermixing.

sequent filter battery runs were made using 1.84 mm coal on top of 0.55 mm sand because:

- The finer sand produced slightly better effluent quality.
- The filter required a lower backwash rate.
- The run lengths were not seriously affected by the use of the finer sand.

Figure 29 shows the effect of filtration rate on the quality of effluent vs. depth in the dual media filter after 11 hours of filtration at two nearly similar influent solids levels. In both cases, the largest part of the solids removal takes place in the coal. The higher filtration rates serves to distribute the solids more uniformly in the coal media, thus improving hydraulic performance at the higher rates. In both cases, the lower rate gave slightly better effluent quality, and all of the improvement took place in or above the coal-sand intermix zone. The data indicate little quality improvement would be obtained by increasing sand media depth. The data do indicate, however, some increased depth of coal would be beneficial at the 6 gpm/sq.ft. filtration rate. Further quality improvement with little reduction in run length undoubtedly could be obtained if a still finer garnet sand were located beneath the silica sand. It does not appear necessary, however, in this application.

Figure 30 shows the head loss vs. time curve generated using 1.84 mm coal - 0.55 mm sand filters at filtration rates of 2, 4, and 6 gpm/sq.ft. when the wastewater contained 12.5 mg/l of suspended solids. The figure indicates that reasonable run lengths become possible using this filter media. Fig. 31 shows the wastewater suspended solids level changes (influent and effluent) during the run. Several observations may be made:

- Effluent suspended solids levels from all filters were generally below 2.0 mg/l and improved with time of filtration.

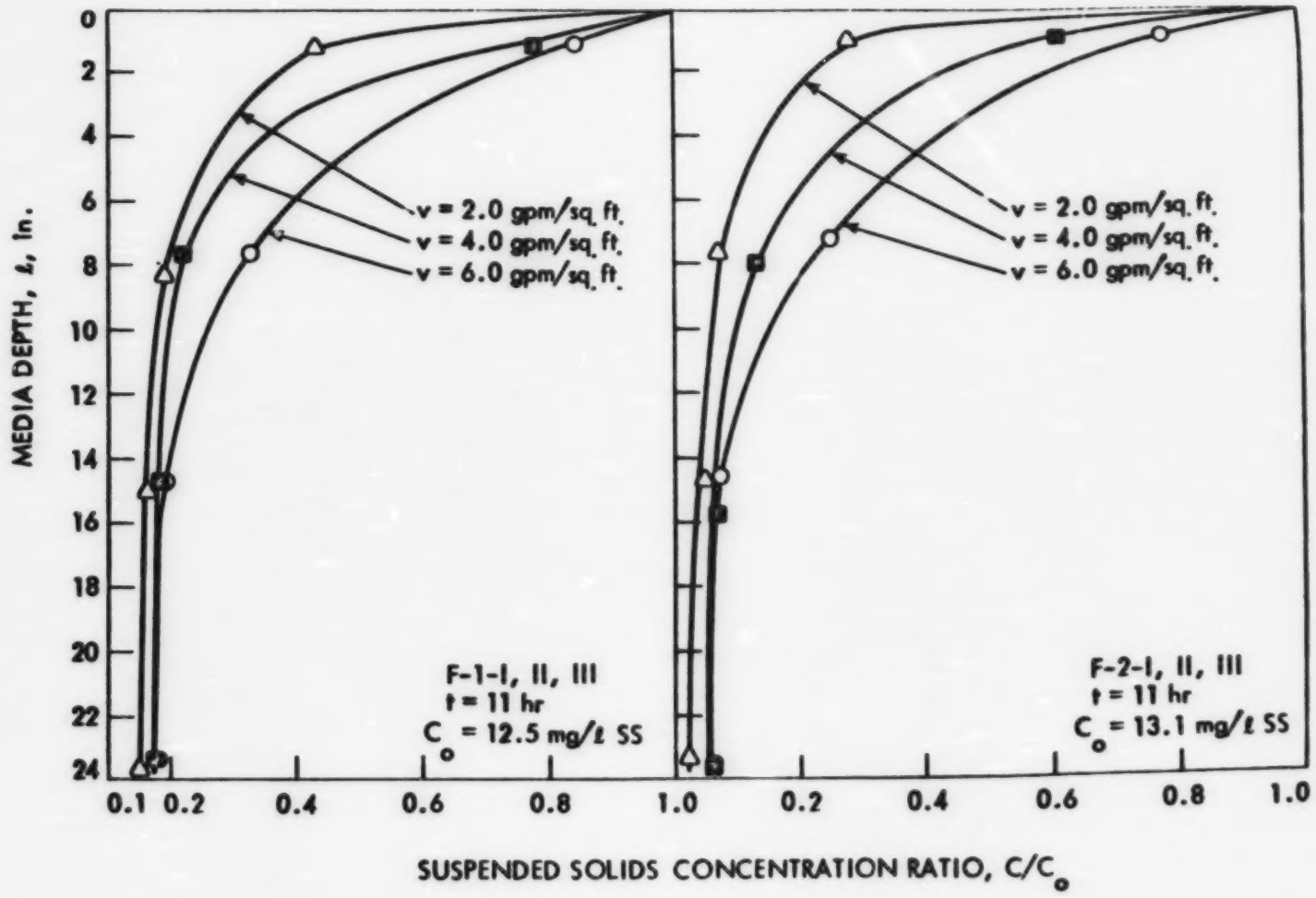


Fig. 29. Comparison of removal efficiencies of various flow rates (12 in. - 1.84 mm anthracite on top of 12 in. - 0.55 mm sand)

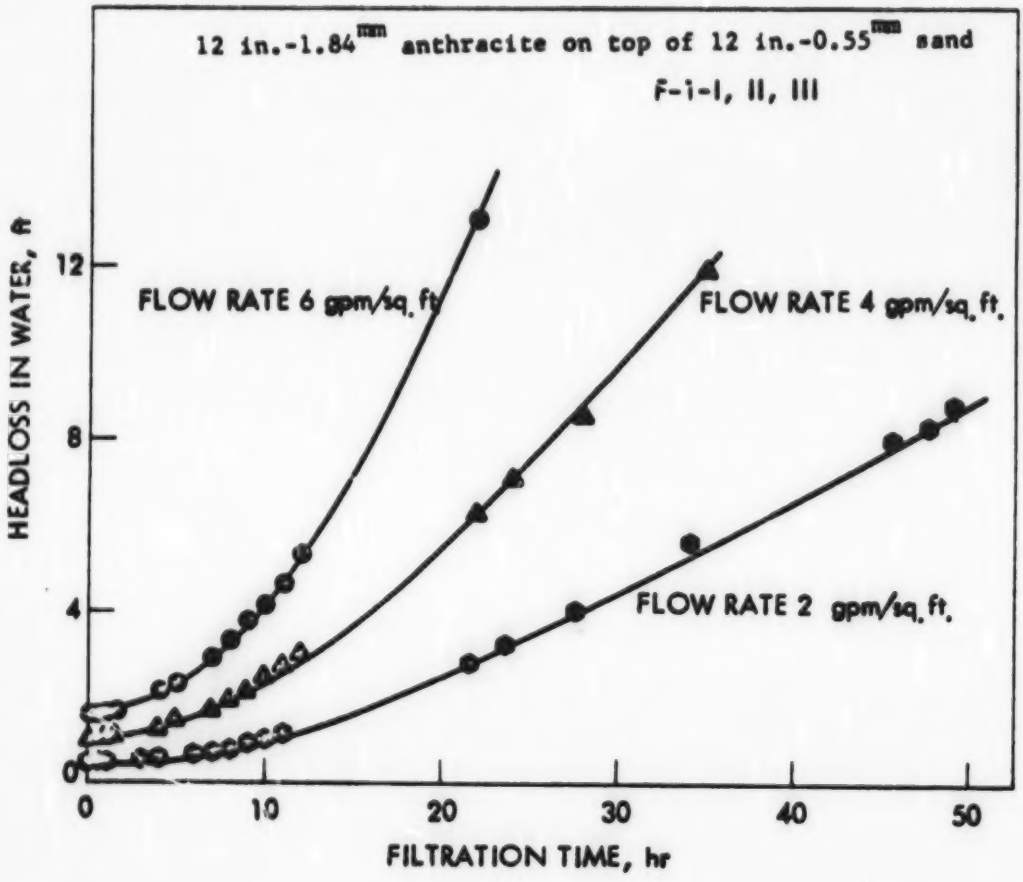


Fig. 30. Comparison of headloss vs. time at various flow rates in Phase F run

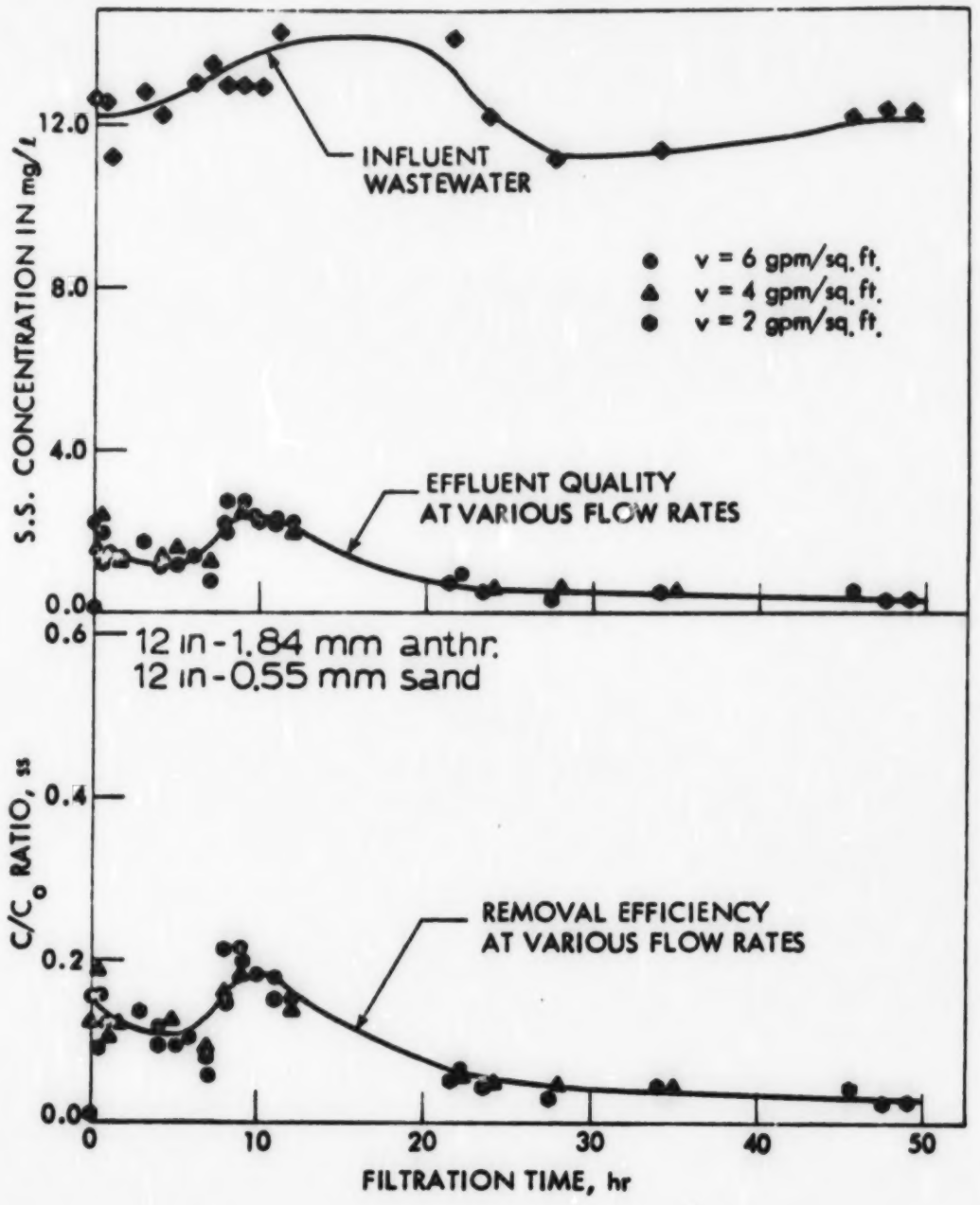


Fig. 31. Comparison of effluent qualities and removal efficiencies at various flow rates, F-1

- The effluent quality at all filtration rates was similar, and
- The effluent quality was more a function of influent suspended solids level than it was a function of filtration rate.

Data from a number of runs similar to that shown in Figs. 30 and 31 were used to prepare the plot of run length vs. SS concentration at various filtration rates shown in Fig. 14. This represents the ultimate relationship which must be generated by pilot plant experience if we are to design technically feasible, operationally effective, economic filters for tertiary wastewater filtration. Until data like these are collected using filters to clarify phosphate precipitation - settled effluents, we will have to continue design of the tertiary filters on a by guess and by gosh basis.

For the removal of residual SS from a trickling filter plant effluent at Ames, Iowa, the above pilot plant study has led to adoption of the following filter plant design criteria:

- Dual Media, 15 inches of 1.84 mm coal and 15 inches of 0.55 mm silica sand.
- Terminal head--10 ft. of water
- Filtration rate--6.25 gpm/sq.ft. at peak 4-hr. rate.
- One, 4-cell variable declining rate filter
- Total filter area, 2940 sq.ft. [18.1 MGD]
- Expected capital cost: $300,000.

SCALE UP FACTORS

Experience to date with filtration indicates that pilot-plant filtration results can be extrapolated directly to any size plant desired. Pilot plants themselves, however, require that the diameter of a cylindrical pilot filter be at least 50 times the diameter of the largest media size. Thus, the minimum pilot filter should have a diameter of at least (3mm anthrafilt x 50 = 150mm or 150/25.4 - 5.9 inches) 6 inches. The effects of media size and depth, flow rate, head losses, and filtering efficiency of a 6-inch diameter filter will be the same as on an 850 sq. ft. filter, the recommended maximum size filter. Thus, scale up is not a problem in filtration.

PROCESS DEPENDABILITY

The process of filtration is one which is unusual in that it will never, if properly designed, provide a grossly inferior quality of effluent. In general, if a filter can provide the desired effluent quality under normal conditions, upsets in pretreatment processes will provide shorter filter runs and not significantly poorer effluent quality. Thus, if under normal conditions the effluent SS are running at 18 mg/l and suddenly increase to a level of 30 or 40 mg/l, the principal effect will be a significant decrease in run length but a relatively lesser increase in effluent SS.

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